PB-214 162

LOW-SULFUR CHAR AS A CO-PRODUCT IN COAL GASIFICATION

G. P. Curran, et al

Consolidated Coal Company

Prepared for:

Environmental Protection Agency

October 1972

DISTRIBUTED BY:

NTIS

National Technical Information Service
U. S. DEPARTMENT OF COMMERCE
5285 Port Royal Road, Springfield Va. 22151

PB 214 162

EPA-R2-72-060
OCTOBER 1972

Environmental Protection Technology

Low-Sulfur Char as a Co-product in Coal Gasification

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. Department of Commerce
Springfield VA 22151

Office of Research and Monitoring
U.S. Environmental Protection Agency
Washington, D.C. 20460

10

BIBLIOGRAPHIC DATA SHEET	1. Report No. EPA-R2-72-060	2.	3. Recipient's Accession No.
4. Title and Subtitle: Low-Sulfur Char as a Co-product in Coal Gasification			5. Report Date: October 1972 6.
7. Author(s): G.P. Curran, W.E. Clark, M. Pell, and E. Gorin			8. Performing Organization Rept. No.
9. Performing Organization Name and Address: Consolidation Coal Company, Inc. Research Division Library, Pennsylvania 15129			10. Project/Task/Work Unit No. 11. Contract/Grant No. EHSD 71-15
12. Sponsoring Organization Name and Address: U.S. ENVIRONMENTAL PROTECTION AGENCY Office of Research and Monitoring Washington, D. C. 20460			13. Type of Report & Period Covered 14.

15. Supplementary Notes

16. Abstracts: A feasibility study is presented for the case where low-sulfur char is produced as a co-product with low-sulfur producer gas in a gasification-desulfurization operation with bituminous coal. Calcium carbonate is used as a sulfur acceptor. Experimental data are also presented to support the design feasibility study. These data show that preoxidized coals are extremely responsive to desulfurization under the conditions used in the feasibility study. It is possible by this method to produce low-sulfur chars containing 0.5% or less sulfur content. The economic evaluation shows that it is possible to produce low-sulfur char at a lower Btu cost than would be the case for complete gasification to low-sulfur producer gas. The value of the char, however, as a boiler fuel is less than that of the producer gas itself. It is, therefore, concluded that there is no clear incentive to produce low-sulfur char as a co-product in a gasification-desulfurization operation.

17. Key Words and Document Analysis. 17a. Descriptors

Air pollution
Feasibility
Coal
Bituminous coal
Gasification
Desulfurization
Charcoal
Sulfur compounds
Acceptor materials
Calcium carbonates
Economic analysis

17b. Identifiers/Open-Ended Terms

Char
Co-products

17c. COSATI Field/Group 13B

1a

18. Availability Statement: Unlimited

19. Security Class (This Report) UNCLASSIFIED

20. Security Class (This Page) UNCLASSIFIED

21. No. of Pages ~~64~~ 70

22. Price ~~$3.00~~ $0.95

EPA-R2-72-060

LOW—SULFUR CHAR AS A CO—PRODUCT IN COAL GASIFICATION

By
G.P. Curran, W.E. Clark,
Melvyn Pell, and Everett Gorin
Consolidation Coal Company, Inc.
Research Division
Library, Pennsylvania 15129

Contract No. EHSD 71-15
Program Element No: 1A2013

Project Officer: D. Bruce Henschel
Control Systems Division
National Environmental Research Center
Research Triangle Park, North Carolina 27711

Prepared for
OFFICE OF RESEARCH AND MONITORING
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C. 20460

ib

October 1972

EPA REVIEW NOTICE

This report has been reviewed by the Environmental Protection Agency and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

ABSTRACT

A feasibility study is presented for the case where low-sulfur char is produced as a co-product with low-sulfur producer gas in a gasification-desulfurization operation with bituminous coal. Calcium carbonate is used as a sulfur acceptor. Experimental data are also presented to support the design feasibility study. These data show that preoxidized coals are extremely responsive to desulfurization under the conditions used in the feasibility study. It is possible by this method to produce low-sulfur chars containing 0.5% or less sulfur content. The economic evaluation shows that it is possible to produce low-sulfur char at a lower Btu cost than would be the case for complete gasification to low-sulfur producer gas. The value of the char, however, as a boiler fuel is less than that of the producer gas itself. It is, therefore, concluded that there is no clear incentive to produce low-sulfur char as a co-product in a gasification-desulfurization operation.

ACKNOWLEDGMENT

The principal investigator in this study was Everett Gorin for Consolidation Coal Company and D. Bruce Henschel was Project Officer for the Environmental Protection Agency in monitoring this contract. Appreciation is expressed to Bruce Henschel for his many helpful suggestions which were of considerable assistance in fulfilling the goals of the research work under this contract.

TABLE OF CONTENTS

		Page
I.	SUMMARY	1
	Table I - Preliminary Economic Comparison of Methods of Producing Low-Sulfur Boiler Fuel	2
II.	CONCLUSIONS AND RECOMMENDATIONS	3
III.	INTRODUCTION	4
IV.	PROCESS DEFINITION	5
	A. General Description	5
	Figure 1 - Flow Diagram for Modified Process to Produce Desulfurized Char Co-Product	6
	B. Design Basis	7
	C. Survey of Process Variables	9
	D. Screening Evaluation - Results and Discussion	10
	Table II - Variable Study for Desulfurized Char Co-Product Processes	12
	E. "Minimum" Gasification Case - Heat and Material Balance	13
	Table III - Further Comparison of Cases 1 and 2	14
	F. High Gasification Case - Heat and Material Balance	15
V.	SUPPORTING EXPERIMENTAL DATA	16
	A. General Background	16
	B. Experimental	16
	C. Results	17
	Figure 2 - Schematic Drawing of Experimental Apparatus	18
	Figure 3 - Differential Desulfurization of Preoxidized Coals - Comparison with Data on LTC Chars	19
	Figure 4 - Differential Desulfurization of Preoxidized Coal at 1500°F	21
VI.	ECONOMICS	22
	Table IV - Product Distribution	23
	Table V-A - Investment, Manpower and Utility Summary Minimum Gasification Case	24
	Table V-B - Investment, Manpower and Utility Summary High Gasification Case	25
	Table VI - Expander Output, Power Output, and Air Compressor Requirement	26

TABLE OF CONTENTS - Cont'd.

		Page
	Table VII-A - Direct Operating Costs, Minimum Gasification Case	27
	Table VII-B - Direct Operating Costs, High Gasification Case	28
	Figure 5 - Cost of Fuel Gas as a Function of the Price Received for Char/Coal at 40¢/MM Btu	29
VII.	BIBLIOGRAPHY	31
VIII.	APPENDICES	32
	APPENDIX A	32
	Figure A-1 - Schematic Flow Diagram - Low-Sulfur Fuels - CO_2 Acceptor Process - Minimum Gasification Case, Dwg. XF-3255	33
	Tables	
A-I	Mass and Heat Balance - Preoxidizer - Gasifier	34
A-II	Mass and Heat Balance - Reactor (Squires)	35
A-III	Mass and Heat Balance - Heat Exchanger C-201	36
A-IV	Mass and Heat Balance - Heat Exchanger C-202	37
A-V	Mass and Heat Balance - Compressor JC-203	38
A-VI	Mass and Heat Balance - Heat Exchanger C-301	39
A-VII	Mass and Heat Balance - Product Gas at 206 psia	40
A-VIII	Mass and Heat Balance - Product Gas at 25.7 psia	41
A-IX	Mass Balance - CO_2 Removal System	42
A-X	Mass Balance - Sulfur Recovery	43
	APPENDIX B	44
	Figure B-1 - Schematic Flow Diagram - Low-Sulfur Fuels - CO_2 Acceptor Process - High Gasification Case, Dwg. XF-3266	45
	Tables	
B-I	Mass and Heat Balance - Preoxidizer - Gasifier	46
B-II	Mass and Heat Balance - Reactor (Squires)	47
B-III	Mass and Heat Balance - Heat Exchanger C-201	48
B-IV	Mass and Heat Balance - Heat Exchanger C-205	49
B-V	Mass and Heat Balance - Heat Exchangers C-203 & C-204	50
B-VI	Mass and Heat Balance - Heat Exchanger C-202	51
B-VII	Mass and Heat Balance - Heat Exchanger C-206	52
B-VIII	Mass and Heat Balance - Heat Exchanger C-207	53
B-IX	Mass and Heat Balance - Compressor JC-203	54
B-X	Mass and Heat Balance - Heat Exchanger C-302	55
B-XI	Mass and Heat Balance - Product Gas at 206 psia	56
B-XII	Mass and Heat Balance - Product Gas at 25.7 psia	57
B-XIII	Mass Balance - CO_2 Removal System	58
B-XIV	Mass Balance - Sulfur Recovery	59

I. SUMMARY

The results of a feasibility study are presented here for the case where low-sulfur char is produced as a co-product with low-sulfur producer gas from bituminous coal in a gasification-desulfurization operation in which $CaCO_3$ is used as a sulfur acceptor.

The results of this study with costs escalated to 1976 and with coal at 40¢/MM Btu are summarized in Table I. Where the total gas and char product are used as fuel to a conventional boiler, the cost of the desulfurized fuel increases with increasing extent of gasification. The minimum incremental cost of the desulfurized fuel to the power plant is 20¢/MM Btu above the cost of coal to the process.

If, instead, the gas component of the product mix is assumed to have a higher value than the char, equal to the value for the total gasification case, i.e., of 70.4¢/MM Btu, the lowest cost of low-sulfur char is still achieved with minimum gasification. In this case, the cost of the low-sulfur char is 11¢/MM Btu greater than that of the coal feed.

Experimental data are presented which show that very low-sulfur levels in the product chars of the order of 0.5% or less may be expected in application of the process considered here. Preoxidized coals are shown to be more readily desulfurized than other carbonaceous materials.

Experimental data indicate that a "flash" desulfurization process using fine preoxidized coal may be feasible. Such a process would be lower in cost than the processes considered, but the magnitude of the cost reduction has not been estimated.

TABLE I

Preliminary Economic Comparison of
Methods of Producing Low-Sulfur Boiler Fuel
Total Fuel Burned in Conventional Boiler

Case	This Study, Char as Co-Product: Minimum Gasification	This Study, Char as Co-Product: High Gasification	Annual Report to OAP (Ref. 1): Total Gasification (Case II)
Plant Investment, $ MM(1)	62.9	109.7	118.0
Annual Costs, $ MM/yr (6132 hr/yr)			
Direct Operating Costs, ex. Coal	6.74	8.89	10.42
Coal at 40¢/MM Btu	33.89	33.89	33.89
Capital Charges at 15%	9.44	16.45	17.70
Less Sulfur Credit at 15 $/L Ton	-1.71	-1.77	-1.83
Less Power Credit at 9 mills/KWH	-1.90	-5.07	-10.65
Net Annual Costs	46.46	52.39	49.53
Fuel Product			
HHV of Gas + HHV of Char, MM Btu/hr	12,338	11,509	10,415
Same, + Sensible Heat in Gas	12,663	12,358	11,480
% of Total as Gas	46.7	73.8	100.0
Fuel Cost, Delivered to Power Station(3)			
¢/MM Btu (HHV)	61.4	74.3	77.6
¢/MM Btu (HHV + Sensible Heat)	60.0	69.1	70.4
MW Excess Power	34.5	91.7	193.0
Efficiency			
Cold	.893	.833	.754
Overall(2)	.925	.917	.879

(1) 1976 operation. Includes escalation and interest during construction at 7-1/2%/yr.
(2) Overall efficiency = [HHV (gas + char) + gas sensible + 3413 (KW x's power)]/HHV in with coal.
(3) Gas and Char assumed to have the same value.

II. CONCLUSIONS AND RECOMMENDATIONS

It is more economical in producing low-sulfur boiler fuel to minimize the extent of gasification in the desulfurizing process. It is not clear, however, that the magnitude of the cost reduction, i.e., ca. 10¢/MM Btu is sufficient to compensate for the disadvantages of producing the low-sulfur char co-product. The disadvantages are: ash handling and removal facilities must be provided at the boiler plant, and the substantial control over NO_x emissions is lost as compared with combustion of an all-gaseous fuel. In addition, the possibility of using the gas as a premium fuel in a combined cycle power plant is largely lost because, if the char is to be burned in a conventional boiler, the gas will have to be fired simultaneously to ensure stable combustion. Also, for the above reasons, the process in which char is a co-product shows no clear incentive over flue gas scrubbing. Total gasification is more expensive, but it can be combined with cleaner combustion and more efficient power generation techniques.

The major incentive for a char co-product process would come about if fluid bed combustion were a developed process. The low-sulfur char could then be used as fuel without need for supplementary gas for firing and without further sulfur control in such a plant. The gas could then be used as fuel for a combined cycle type power plant.

Some cost savings over the cases considered here would be possible by use of a modified "flash" desulfurization process. It is recommended that experimental work be conducted in Consol's continuous bench-scale unit to establish the feasibility of such a process.

III. INTRODUCTION

A study[1] had been prepared for the Control Systems Division of the EPA on the adaptation of the CO_2 Acceptor Process to the problem of producing clean low Btu producer gas from high-sulfur bituminous coals. The study included a preliminary experimental evaluation of the process concept as well as an economic feasibility study. The feasibility study was directed solely to the case where complete gasification of the coal feed was effected. This was done because it was felt desirable not only to eliminate sulfur but also ash from the fuel product. Another reason was that subsequent combustion of an all gaseous fuel would be expected to yield a lower and more controlled level of NO_x emission.

The results of this study as well as some associated experimental work are given in the Annual Report[1] to the Control Systems Division of EPA. This report will simply be referred to hereafter as the Annual Report.

It was pointed out in the Annual Report[1] that prior Consol experimental data had shown that the process could easily be reoriented to produce low-sulfur char as a co-product with the low-Btu gas. This fact was reconfirmed by experimental data obtained during the aforementioned study. It was quite clear also, that the total low-sulfur fuel mix could be produced at a lower cost where char is a co-product.

Accordingly, Consol Research, was requested by the Control Systems Division to revise the previous feasibility study to include the case where low-sulfur char is a co-product. This has now been done and the results are reported herein.

A similar study[2] had previously been prepared for the Office of Coal Research. In this instance, a feasibility study was prepared wherein low-sulfur char was considered as a co-product with production of high Btu pipeline gas.

Some new experimental data are also briefly presented which demonstrate the ease of desulfurization of preoxidized coals at very mild conditions.

IV. PROCESS DEFINITION

A. General Description

The sulfur recovery system used in this process has been patterned after that of Case II in the Annual Report.[1] This is because experimental work carried out subsequent to the preparation of the Annual Report has shown that the sulfur recovery system of Case I, though somewhat cheaper than that of Case II, is of doubtful technical feasibility.

The process considered here differs from that discussed in the Annual Report[1] in that with low-sulfur char as a co-product the need for a CO_2 acceptor regenerator ($CaCO_3 \rightarrow CaO$) as the ultimate sink for ungasified char disappears. The heat, formerly supplied by the acceptor action in the gasifier is now supplied by additional combustion of air. In all the cases discussed here there is no CO_2 acceptor action, only H_2S acceptance. Sulfur is removed in the gasifier by the reaction,

$$CaCO_3 + H_2S = CaS + CO_2 + H_2O.$$

The sulfided acceptor is regenerated by reversing the above reaction in a separate vessel exactly as described in Case II of the Annual Report.[1] This regeneration reaction is referred to elsewhere in this report as the "Squires" reaction. In all cases, gasifier conditions are such that an adequate CO_2 partial pressure exists to prevent calcination of $CaCO_3$ entering the gasifier.

1. Base Case

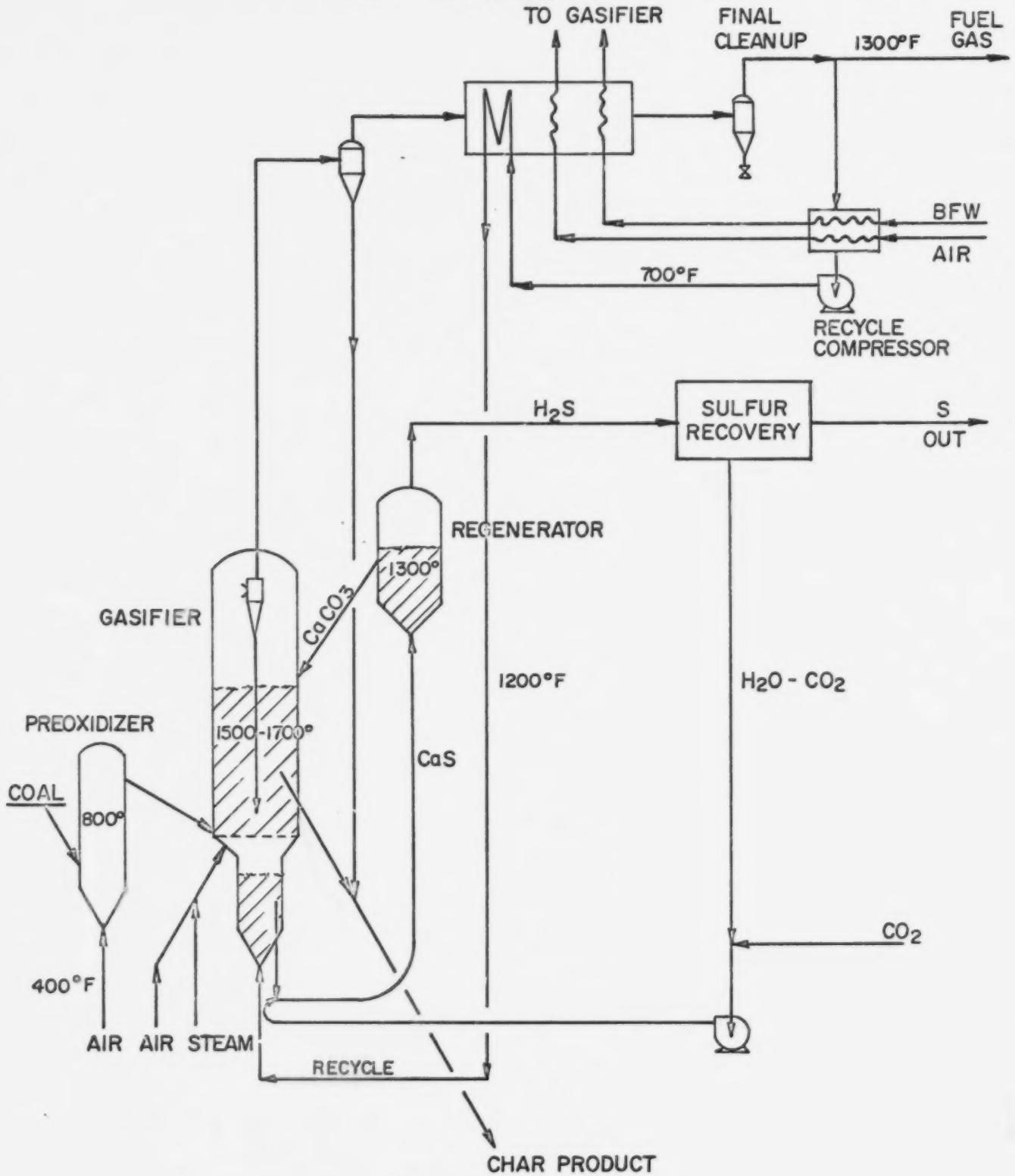
A simplified flow diagram is shown in Figure 1. After preoxidation at adiabatic conditions, the coal and its preoxidation products are fed directly to the bottom of the gasifier vessel along with additional air and steam which have been preheated by exchange with the gasifier offgas. All the heat duty for the gasifier is supplied by partial combustion of a portion of the coal with air.

Regenerated acceptor ($CaCO_3$ + MgO) enters at the top of the gasifier, showers down through the fluidized bed of char, and segregates in a "boot" below the entry points of the coal, air and steam. The acceptor in the boot is fluidized by a stream of recycled product gas. The recycle gas is cooled by exchange with the incoming air and boiler feed water, recompressed, and then is reheated by exchange with the gasifier offgas.

All of the recirculating acceptor is picked up from the gasifier boot by a stream of CO_2-steam and is carried to the regenerator vessel. The regenerated acceptor then is returned to the gasifier. The H_2S-bearing stream from the regenerator is converted to elemental sulfur by the Wackenroder reaction (liquid-phase Claus) as described on page 27 of the Annual Report.[1]

Figure 1

FLOW DIAGRAM FOR MODIFIED PROCESS TO PRODUCE DESULFURIZED CHAR CO - PRODUCT



2. Improved System - Use of Highly Caking Pittsburgh Seam Coal

The base case gasifier-desulfurizer system requires the use of a relatively coarse coal feed to achieve a reasonable vessel cross-sectional area. This is true because, the char is the continuous phase in the gasifier-desulfurizer and its fluidization properties, i.e., average size and particle density must be compatible with the fluidization velocity of 1.09 ft/sec used in the design cases discussed below.

Previous work presented in the Annual Report[1] has shown that adiabatic preoxidation of the highly caking Pittsburgh Seam coals is not effective at elevated system pressure in producing an operable gasifier feed when relatively coarse (24 x 100 mesh) coal is used. Evidence is at hand, however, that finer sized coal (say 65 mesh x 0) would be operable at adiabatic preoxidation conditions.

For production of a desulfurized char co-product, the relatively coarse particle size required to form a dense-phase fluidized bed is no longer needed since gasification of fixed carbon and hence long residence time is not required. In this improved case the acceptor would be the continuous phase in the gasifier, fluidized at 2-3 ft/sec. Finely sized char would be fed at the bottom of the bed, and would pass up through it. By controlling the acceptor bed expansion, the mean retention time of the finely sized char particles can be held in the range of 10 to 20 minutes. The char particles ultimately would be elutriated from the bed of coarse acceptor and recovered as product.

Thus, the improved case provides not only a more practical preoxidation system than the base case for processing of Pittsburgh Seam coals, but also has the additional advantage of reduced vessel costs.

At the time that this study was carried out, experimental data were not at hand to show that the desired degree of desulfurization could be achieved at the low residence times required for the improved case. Such data have since been obtained which show that this is indeed feasible as will be reported below.

B. Design Basis

A screening type evaluation of twelve specific cases around the flow sheet of Figure 1 was made to select two specific cases for economic evaluation. The computer program described in the Annual Report was used, with suitable changes to fit the modified system of Figure 1, for this preliminary evaluation. The major process restraints, i.e., those related to the fluidization properties of the materials handled, the thermodynamics and kinetics of the acceptor reactions, and of the carbon steam reactions, etc., remain the same as those set forth previously.[1]

Also, the design of the sulfur recovery section utilizing the "Wackenroder reaction" is based entirely on the Case II system described in the Annual Report.

The preliminary screening evaluation is aimed at determining the overall heat and material balance, vessel sizes and process efficiency as a function of the selected process variables for each case.

The following design assumptions are common to all the cases considered:

1. Coal

The same coal composition was used as in the original feasibility study, i.e., high-sulfur eastern bituminous coal having 4.3% sulfur content. The analysis is duplicated below for the convenience of the reader.

Feed Coal Analysis

Moisture (as received)	6.0 wt %
Ultimate Analysis, Wt %	
Hydrogen	4.8
Carbon	69.8
Nitrogen	1.2
Oxygen	7.6
Sulfur	4.3
Ash	12.3
Higher Heating Value	12,700 Btu/lb (MF basis)

2. Char Sulfur Content

The sulfur content of the product char, for the purpose of this work, was conservatively taken as 0.75 wt % regardless of the level of gasification of fixed carbon. This assumption is based on previous Consol work with low-temperature carbonized (LTC) char.

Work carried out since this design study was made shows, however, that preoxidized coals desulfurize more readily than LTC char, and sulfur contents well below 0.75 wt % should be achievable. These data as well as other background data are discussed in Section V of this report.

3. Preoxidation Level

The extent of preoxidation* was maintained within the adiabatic restraints of the process, and as discussed previously,[1] amounts to about 8.6% for a preoxidizer operated at 800°F. It was shown experimentally,[1] that preoxidation to this level, was sufficient to render 24 x 100 mesh Illinois No. 6 coal operable in the gasifier. However, when high sulfur Pittsburgh Seam coals of the same size consist are used, it has been shown that the extent of preoxidation required to establish operability is in excess of 20%. Thus, with coarse coals of this type, a process modification is required wherein heat is removed from the preoxidizer. The alternative is to use finer coal and the improved process as discussed above.

The economic analysis for the base case, thus strictly applies only to the use of the more weakly caking coals such as Illinois No. 6. The coal analysis used in this study, however, actually is that of a high-sulfur Pittsburgh Seam coal. It is felt, however, that the overall heat and material balance would not change significantly if the substitution with Illinois No. 6 coal were made.

---

* Extent of preoxidation is expressed as a percentage and is defined by 100 x lbs O_2 consumed per lb of moisture-free coal.

4. Acceptor Sulfur Reactions

It is essential, in order to achieve the required char sulfur levels, that rapid and efficient removal of H_2S inhibitor be effected in situ by means of the reaction,

$$CaCO_3 + H_2S = CaS + H_2O + CO_2.$$

Prior laboratory data by Ruth, et al.,[3] have demonstrated that this reaction is very rapid. It was also demonstrated in the operation of our continuous unit[9] that efficient H_2S removal was effected in the gasifier via the above reaction.

5. Product Gas Cleanup

The assumption was made that particulate matter and alkali in the product fuel gas, after cooling to 1300°F, could be reduced sufficiently by high-pressure drop cyclones to allow sustained operation of gas turbine engines or expanders. The product gas is cooled to 1300°F in all cases by heat exhange with the incoming air, steam and recycle gas streams as shown in Figure 1.

6. Preheating of Inlet Streams

In the cases where the gasifier temperature is 1600°F or higher, the maximum preheat temperature was taken as 1200°F in order to avoid use of high alloy tubing in the heat exchangers. In these cases, the recycle gas and steam were preheated to 1200°F. The remaining available heat was used to preheat the gasifier air stream.

For the 1500°F cases, the preheat temperature limit was raised to 1350°F in order to keep the maximum tube wall temperature roughly comparable with that of the higher temperature cases.

In all the cases, the preoxidizer air stream temperature was that at the compressor outlet (398°F for the 15 atm cases).

7. Acceptor Make-up Rate

As in the previous feasibility study the acceptor make-up rate was 0.5%, i.e., 0.005 mols fresh $MgCO_3 \cdot CaCO_3$ added per mol $MgO \cdot CaCO_3$ circulated through the gasifier. No data are available as to whether this make-up rate would be sufficient to maintain adequate acceptor activity. However, the acceptor circulation rate is only about 1/10 that of the previous cases in which CO_2 acceptor action was used. Thus, the incremental cost of a higher make-up rate would be relatively small.

C. Survey of Process Variables

The specific conditions discussed here form the basis for the twelve cases which were considered in the primary evaluation.

1. Percent of Fixed Carbon Gasified

By varying the gasifier temperature and the amount of inlet steam, a wide range of fixed carbon gasification can be achieved which in turn alters the ratio of energy available in the gas and char co-product streams.

Temperatures below 1500°F were not considered since our OCR work has shown that this is the minimum temperature at which substantially complete hydrocracking of tarry matter occurs. Temperatures above 1700°F were not used, primarily to avoid the problem of ash slagging at the gasifier air inlet point. Also at temperatures above 1700°F, in some situations the CO_2 partial pressure would become insufficient to prevent calcination of $CaCO_3$.

2. Extent of Conversion of $CaCO_3$ to CaS

As noted above, no data on the effect of process cycling on acceptor activity exist. For the purpose of this study, a base activity of 0.20 was used, i.e., 20 mols of CaS are formed per 100 mols total calcium on each passage through the gasifier.

3. Regenerator Temperature

The base temperature for the acceptor regenerator ("Squires" reactor) was taken as 1300°F, the same as in the original feasibility study.

4. System Pressure

To be compatible with cases in the Annual Report,[1] the base system pressure was taken as 15 atm (206 psia).

5. Fluidization of the Gasifier Boot

Air and steam cannot be used to fluidize the acceptor in the gasifier boot because the CaS would be oxidized to $CaSO_4$, making regeneration impossible via the "Squires" reaction. Therefore, recycle gas must be used. The base amount of recycle gas corresponds to the ratio of boot cross-sectional area to that of the gasifier of 1/6.

D. Screening Evaluation - Results and Discussion

The scope of this study does not include specification of how the co-products, low-sulfur fuel gas and char, are to be used ultimately in the generation of clean power. Nevertheless, some conceptual assumptions had to be made to allow comparisons of the merits of the various cases.

Both fuels can be burned in a conventional boiler, with suitable burner modifications. In this instance, the product gas is passed through expanders to reduce the pressure to about 25 psia. The expanded gas is delivered at about 700°F to the burners. The expanders drive compressors for the process air and produce additional power which is sold. The desulfurized char is cooled, depressured and delivered to the station silos.

The low-volatile char cannot be burned in a conventional boiler without supplemental firing. By simultaneously burning the fuel gas co-product stable combustion can be assured, even for the cases which have the minimum fuel gas/char ratio, since the gas supplies about 45% of the heating value of the total fuel.

For comparison among the various cases, an overall efficiency was defined as:

$$\text{Overall efficiency} = \frac{[\text{HHV (gas + char)} + \text{gas sensible heat} + 3413\ (\text{KW excess power})]}{\text{HHV in with coal}}$$

The excess power is that available for sale after subtracting the power required for compression of air and recycle gas and for miscellaneous drives from the gross power output of the expander.

Another method of utilization is to burn the gas in a combined cycle power plant. For the purpose of this study, overall station heat rates were calculated for the supercharged boiler combined cycle described on page 54 of the Annual Report.[1] In these instances, the desulfurized char co-product would have to be burned in a fluidized bed boiler since supplemental firing is not possible. For the char portion of the cycle, a thermal efficiency of 38%, based on the HHV of the char was assumed, without specifying the exact nature of the cycle.

Detailed heat and material balances were calculated for twelve cases. Some of the pertinent results of the calculations are summarized in Table II which also shows the same data for Case II in the original feasibility study.[1]

Cases 1 through 7 show the impact of the extent of fixed carbon gasification over the range of 0 to 65%. Gasifier temperature, amount of inlet steam, and the gasification rate all are compatible with our available data on gasification kinetics.

All the cases having more than zero percent fixed carbon gasified show a decreased overall efficiency (conventional cycle) although the range is not large. For the combined cycle, the overall station rates improve modestly with increasing extent of fixed carbon gasification. However, when the gasifier cross-sectional areas are compared, it is obvious that the lowest cost per unit of total energy in the co-products occurs at zero percent fixed carbon gasified. Direct operating costs and capital charges in these processes are sensitive to the gasifier vessel volume and to the amount of process air which is required, as study of the Economic Evaluation Section of the original feasibility study presented in the Annual Report[1] will show.

Case 10 shows that by roughly doubling the acceptor circulation rate only a slight penalty is involved in either efficiency or in vessel volume. Case 11 shows that decreasing the regenerator temperature to 1200°F causes a nearly negligible penalty in efficiency and vessel volume. If subsequent work shows that the kinetics of the "Squires" reaction are adequate at 1200°F, the total cost of the plant probably will be decreased because the sulfur recovery section investment will be considerably lower.

TABLE II

Variable Study for Desulfurized Char Co-Product Processes

0.75% Sulfur in Product Char.
15 atm System Pressure Unless Noted.

Basis: 100 lb dry coal fed to preoxidizer

Case	1	2	3	4	5	6	7	8(4)	9(1)	10	11	12(6)	Case II
Gasifier													
Temperature, °F	1500	1700	1700	1650	1650	1650	1600	1700	1500	1500	1500	1500	1700
CaS/Σ Ca	←				.20				→	.10	.20	.20	.0187
Regenerator Temperature, °F	←				1300				→		1200	1300	1300
% Fixed C Gasified	0	52	65	48	41	34	20	65	0	0	0	0	65
Mols to Gasifier													
Steam	0	1.970	2.710	2.560	2.113	1.616	1.322	2.506	0	0	0	0	3.92
Air (includes preox. air)	3.417	8.272	9.540	8.339	7.638	6.832	5.616	9.184	3.345	3.636	3.535	3.143	6.39
Recycle	2.802	6.646	7.669	6.849	6.271	5.612	4.741	1.0	0	2.908	2.859	2.627	6.51(3)
Mols Product Gas	6.237	14.793	17.070	15.244	13.958	12.490	10.552	16.521	6.167	6.472	6.364	5.847	23.00
Mols $MgO \cdot CaCO_3$ to Gasifier	.598	.620	.622	.607	.608	.608	.599	.624	.598	1.194	.597	.595	5.878
R_T, Gasification Rate(5)	0	59	75	56	48	39	22	75	0	0	0	0	75
Heating Value of Product Gas, Btu/ft³, wet	216	135	125	124	130	140	150	132	221	205	210	233	115
Gasifier Cross Section, % of Case II	38	100	115	100	92	82	68	81	13	39	39	28	100
PPM H_2S in Product Gas	279	159	195	307	274	232	337	174	272	297	289	378	240
% Sulfur Removed from Coal Feed	89.1	92.5	92.7	90.5	90.6	90.7	89.3	93.1	89.2	89.0	89.1	88.8	96.2
Cold Efficiency	.8933	.8334	.8114	.8219	.8348	.8497	.8649	.8225	.8954	.8870	.8899	.8973	.754
% of Total Product HHV in Gas	45.38	71.89	78.91	68.96	65.56	61.95	55.07	79.19	45.50	44.99	45.17	45.62	100
P_{H_2} at Top of Gasifier, atm	1.67	2.46	2.42	2.41	2.44	2.48	2.40	2.44	1.66	1.67	1.67	1.92	2.23
Air Preheat Temperature, °F	1320	1200	930	750	930	1180	1200	540	1120	1260	1280	1430	398
Recycle Preheat Temperature, °F	1350	1200	1200	1200	1200	1200	1200	1200	--	1350	1350	1350	1800
Btu HHV in Char	619,700	297,520	217,330	324,000	365,770	410,610	493,520	217,330	619,700	619,700	619,700	619,700	0
Btu HHV in Gas	514,830	760,900	813,150	719,810	694,430	668,510	604,900	827,200	517,410	506,800	510,500	519,900	957,580
Overall Station Rate, Btu/KWH for Combined Cycle plus Fluidized Bed Boiler	9060	8815	8810	8930	8920	8900	8950	8700	9030	9120	9095	8990	8570
(4) KW Excess Power	4.53	11.05	12.89	11.73	10.67	9.48	8.16	13.04	4.75	4.60	4.56	4.83	17.75
Btu Sensible Heat in Gas	31,500	66,900	77,600	70,600	64,600	57,600	49,900	74,900	31,200	32,400	32,000	27,300	98,700
Btu HHV + Sensible Heat in Gas	546,330	827,800	890,750	790,410	759,030	726,110	654,800	902,100	548,610	539,200	542,500	547,200	1,056,280
Overall Efficiency *	.930	.916	.907	.909	.914	.920	.926	.916	.933	.925	.927	.933	.878

* Overall efficiency = [HHV (gas + char) + gas sensible + 3413 (KW excess power)]/HHV in with coal.

(1) Acceptor is continuous phase. Gas velocity is 2 x other cases.

(3) System pressure = 20 atm.

(3) Regenerator gas.

(4) D gasifier/D boot = 1/45.

(5) Pounds fixed carbon gasified/pounds fixed carbon in bed/minute x 10^4.

(6) Entire product mix burned in conventional boiler.

The effect of system pressure is indicated by Case 12 which shows that by increasing the system pressure from 15 to 20 atm, modest improvements in efficiency and vessel volume occur. However, the vessel cost will not decrease in direct proportion to the cross-sectional area. Note that the smaller air and recycle gas requirements lead to violation of the 1350°F limit on the air preheat temperature.

Case 8 (to be compared with Case 3) shows that appreciable improvement in efficiency and in vessel costs would occur if the recycle gas flow to the acceptor boot could be reduced considerably. In Case 8 the ratio of boot cross-sectional area to that of the gasifier is about 1/45, compared with a ratio of 1/6 in the other cases. How practical such a drastic reduction in cross-sectional area would be is not known at this time.

Case 9 corresponds to the improved process where fine preoxidized coal is passed through a bed of acceptor. The data of Table I show that such a process would be very attractive from the point of view of investment and operating cost since the vessel volume would be reduced drastically.

Case 1 was chosen for the more detailed economic study to exemplify the minimum gasification case, in place of Case 9, since the experimental data on the feasibility of achieving the sulfur levels required via Case 9 were not available at the time the decision was made.

Because the vessel cross-section is much smaller, Case I clearly will show a lower cost for the total low-sulfur fuel product than the other cases given in Table II where a substantial degree of fixed carbon gasification is effected.

Case 2 was, however, also chosen for economic evaluation to cover the eventuality that a high ratio of fuel gas/char may be desirable in some instances, i.e., for a combined cycle power plant fuel. Case 3 corresponds to an even higher degree of gasification of fixed carbon, but Case 2 was chosen in preference to Case 3 on the basis of lower vessel costs. Case 2 is hereafter referred to as the "high" gasification case. Some further details of Cases 1 and 2 are shown in Table III.

E. "Minimum" Gasification Case - Heat and Material Balance

A schematic, but more detailed flow sheet than Figure 1 for this case is given in the Appendix as Figure A-1. The streams are numerically identified for the purposes of the detailed heat and material balances around the different sections of the plant which are given in Appendix A as Tables A-I through A-VIII, inclusive.

Certain minor changes were made relative to the balances and flow sheet presented in Table II and Figure 1, respectively. In Figure A-1 the gas fludizing the boot is recycled directly at 1300°F without prior cooling as given in Figure 1. If this design proves not to be feasible, the gas may be cooled as shown in Figure 1 prior to recycle compression. The difference in cost of the two procedures is well within the precision of the estimate.

TABLE III

Further Comparison of Cases 1 and 2

Basis: 100 lb dry coal fed to preoxidizer

Case	1	2
Wt % Char Yield	54.9	32.9
H, wt % (dry basis)	0.5	0.4
C	76.3	61.5
N	~ 0	~ 0
O	~ 0	~ 0
S	0.75	0.75
Ash	22.4	37.1
% of Coal HHV in		
Gas	40.5	59.9
Char	48.8	23.4
Cold Efficiency, %	89.3	83.3
Sulfur Distribution, lb		
in with coal	4.30	4.30
out with char	.412	.247
out with gas	.056	.075
out with Acceptor	3.832	3.978
% S Rejected	89.1	92.5

The product gas to the turbine expander is also available now at a lower temperature than given in Figure 1, i.e., 969°F vs. 1300°F and the delivered expanded gas temperature is correspondingly reduced from the previous value of 700°F to 462°F. The main reason for the reduction in temperature is because of the need to blend in with the product gas that portion of the gas from which CO_2 had been removed in the CO_2 recovery system. The CO_2 recovery system operation is required to supply CO_2 to the "Squires" reaction in the sulfur recovery system. Figure 1 is not highly detailed, and the CO_2 recovery system is not included.

F. High Gasification Case - Heat and Material Balance

The flow sheet and material balances for this case are given in Appendix B as Figure B-1 and Tables B-I through B-XIV, respectively. Figure B-1 is essentially a more detailed version of the basic simplified flow sheet of Figure 1. The major differences between the flow sheets for the high and low gasification cases are that (1) in the former case gasifier steam is generated from the offgas and in the latter case no steam is generated or used, and (2) the offgas in the former case is cooled before recycle to the gasifier boot.

In both flow sheets, the details of the sulfur recovery system are not shown. The sulfur recovery flow sheet, however, is identical to the corresponding section of Case II and is given in Figure 7 of the Annual Report.[1]

V. SUPPORTING EXPERIMENTAL DATA

A. General Background

Experimental work carried out some time ago at Consol Research[4] showed that low-temperature carbonized (LTC) chars undergo a remarkable degree of desulfurization simultaneously with gasification by steam-hydrogen mixtures. It was recognized, however, that the active desulfurizing agent was hydrogen and not steam. Data were later published by Consol Research on both the kinetic and equilibrium relationships in the desulfurization of LTC chars[5] by hydrogen alone and devolatilization gases containing hydrogen.

Results obtained in the continuous desulfurization of LTC char were also reported.[5] A fluid bed bench-scale unit[5] was employed in which the use of both once-through hydrogen and recycle devolatilization gases was studied. The extent of desulfurization achieved under these conditions was shown to be limited by equilibrium and not kinetics.

It was recognized that in order to arrive at a satisfactory commercial process it was desirable to use the "natural" hydrogen evolved by devolatilization of the char[6a] on heating to desulfurization temperatures, and it was essential to carry out the desulfurization process in the presence of an H_2S acceptor[6b thru 6f] to eliminate the hydrogen sulfide inhibition of the process.

The effectiveness of the use of a "showering" lime acceptor in desulfurizing a bed of char simultaneously undergoing partial gasification in a continuous fluid bed bench-scale unit has also been reported.[7] Char produced from a Pittsburgh Seam coal was reduced in sulfur to 0.29 wt % by this technique. Similar desulfurization results were also given in work reported in the Annual Report.[1] The organic sulfur content of the char gasification residue was reduced to 7% of that in the feed char.

B. Experimental

The previous work,[5] in which differential desulfurization of LTC char was studied, was conducted by fluidizing a small batch of char with a large excess of H_2, i.e., H_2/char ratios in the range of 140-400 SCFH/lb were used. This was sufficient to substantially eliminate all but minor H_2S inhibition effects.

The present work was carried out with preoxidized coals previously prepared. Details of the preparation are given in the Annual Report.[1]

A summary of the materials used is given below:

Sample No.	Coal Feed to Preoxidizer	Preoxidation, Wt. %	Preoxidation Temp., °F
6P	Illinois No. 6, Hillsboro Mine	8.7	810
4P	Pittsburgh No. 8, Ireland Mine	27.9	750

A somewhat different experimental technique was used in this work. The preoxidized coal was injected into a bed of calcined dolomite fluidized with H_2 at 1 atm pressure. This insured absence of H_2S inhibition.

The reactor and apparatus for injection of preoxidized coal are shown schematically in Figure 2.

The first step in a run was to calcine 35 grams of Tymochtee dolomite (100 x 200 Tyler mesh) in air at 1650°F for 30 minutes. The reactor was quenched and the air was replaced with hydrogen. The 4-way stopcock was cracked so that hydrogen purged the reservoir but no coal came through.

The reactor was then immersed and equilibrated at temperature. Coal feed was started by fully opening the stopcock. It took 30 seconds to one minute to feed the coal charge of 3-5 grams. The timer was started when about half of the coal had been fed.

Using this technique, the bed temperature never dropped more than 100°F and generally returned to temperature within 1-1/2 minutes.

Run 4P product could not be fed with either H_2 or N_2. It formed agglomerates in the feed tube. This material was run by mixing it with the calcined dolomite and then immersing the reactor containing both in the sand bath.

The conversion of CaO to CaS was always less than 10%. A flow of 1.5 SCFH of H_2 was standard. This gave a fluidizing velocity of 0.3-0.4 ft/sec.

E. Results

A comparison of results with preoxidized coals with previous data with LTC chars is given in Figure 3. The weight percent sulfur elimination is defined as follows:

$$\text{Wt \% Sulfur Elimination} = \frac{(\text{gms S in Feed Char} - \text{gms S in Product Char})}{\text{gms S in Feed Char}} \times 100$$

The data points for the preoxidized coals are shown in the graph, while the line is the least squares line drawn under the "unproven" assumption that the extent of sulfur elimination is independent of feedstock.

Figure 2

SCHEMATIC DRAWING OF EXPERIMENTAL APPARATUS

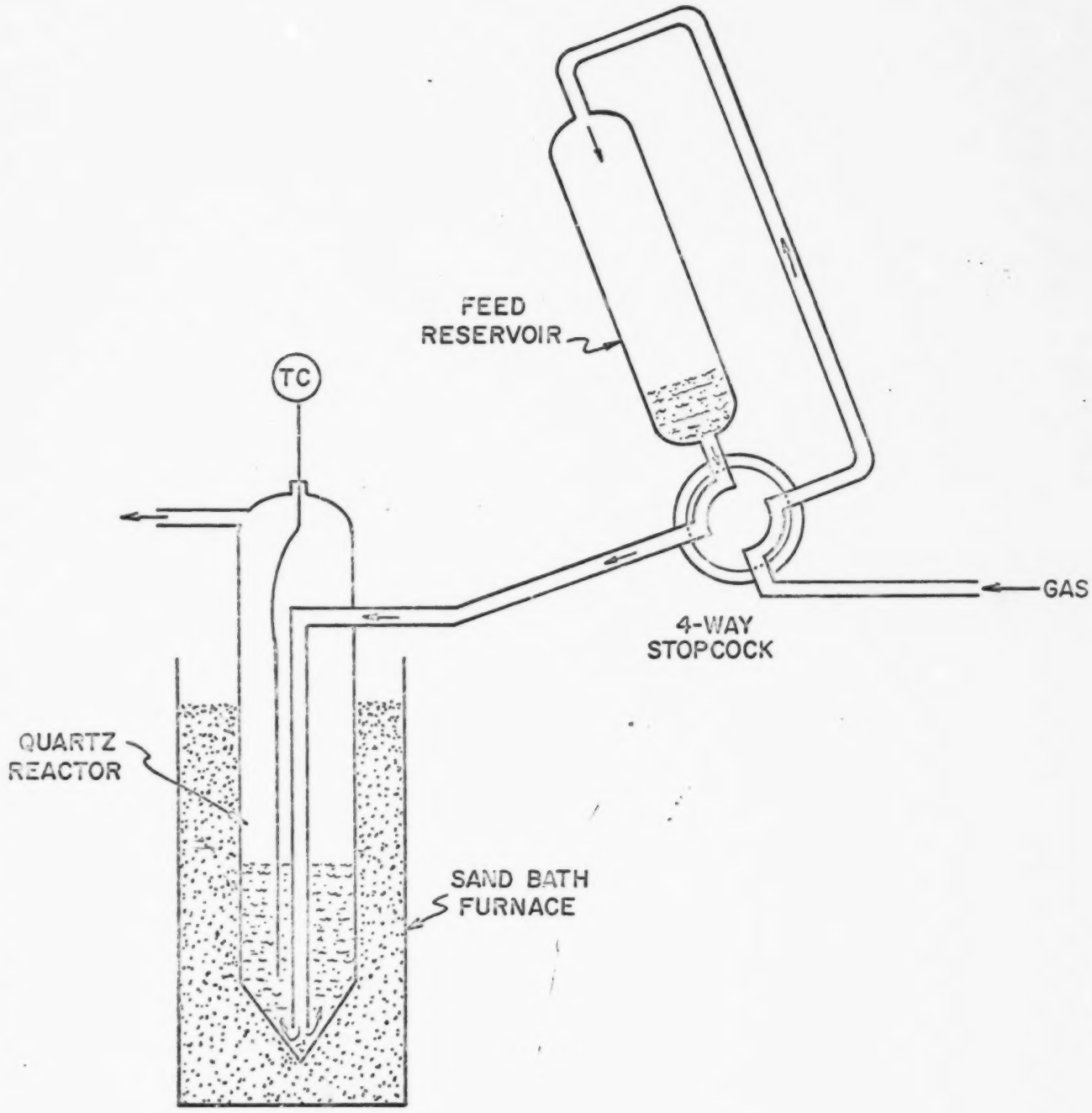
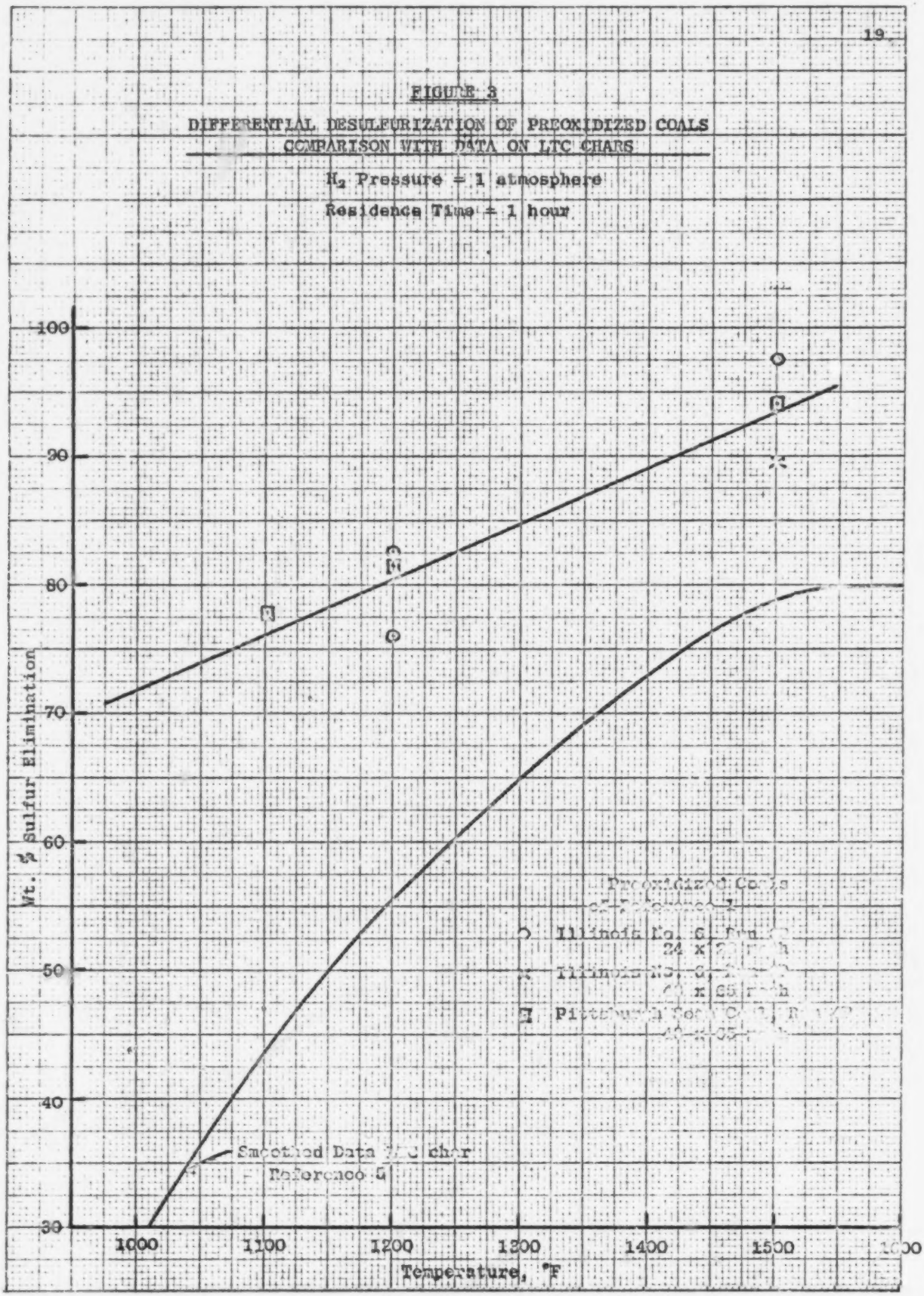


FIGURE 3

DIFFERENTIAL DESULFURIZATION OF PREOXIDIZED COALS
COMPARISON WITH DATA ON LTC CHARS

H_2 Pressure = 1 atmosphere

Residence Time = 1 hour



K&E 10 X 10 TO ½ INCH 46 1323
7 X 10 INCHES
KEUFFEL & ESSER CO

It is immediately obvious that the preoxidized coals desulfurize more readily than the chars, and that the differential becomes more marked as the temperature is reduced.

Figure 4 shows the reduction in sulfur content of preoxidized Pittsburgh Seam coal versus time under process conditions, i.e., at 1500°F. It is noted that the design value of 0.75 wt % sulfur is reached in only 10 minutes residence time.

The char residence time provided in the minimum gasification case cannot be specified exactly since it depends on the average particle size and density of the bed solids which is not known at this time. The order of magnitude is, however, two hours while the average H_2 partial pressure in Case I is slightly higher, i.e., 1.1 atm, than that used in the experimental work. The residence times and the hydrogen partial pressure in the high gasification cases are even higher. It is, therefore, clear that the sulfur content of the product char should be well below the level of 0.75 wt % specified.

As was mentioned above, it is likely that to establish operability with Pittsburgh Seam coals within the framework of adiabatic preoxidation, it will be necessary to use a fine grind of the coal feed. This means that one must resort to the improved case discussed in Section IV-A2 to effect desulfurization. The results shown in Figure 4 suggest that adequate desulfurization can be achieved at the short residence times required by the improved case.

Figure 3 suggests that adequate desulfurization can be accomplished at temperatures well below the 1500°F temperature used in the minimum gasification case. Operating at lower temperatures would produce a higher ratio of char to gas but would likely produce a lower cost net product. It should be pointed out, however, that in this instance one could not use $CaCO_3$ as the acceptor since the equilibrium in the acceptor reaction,

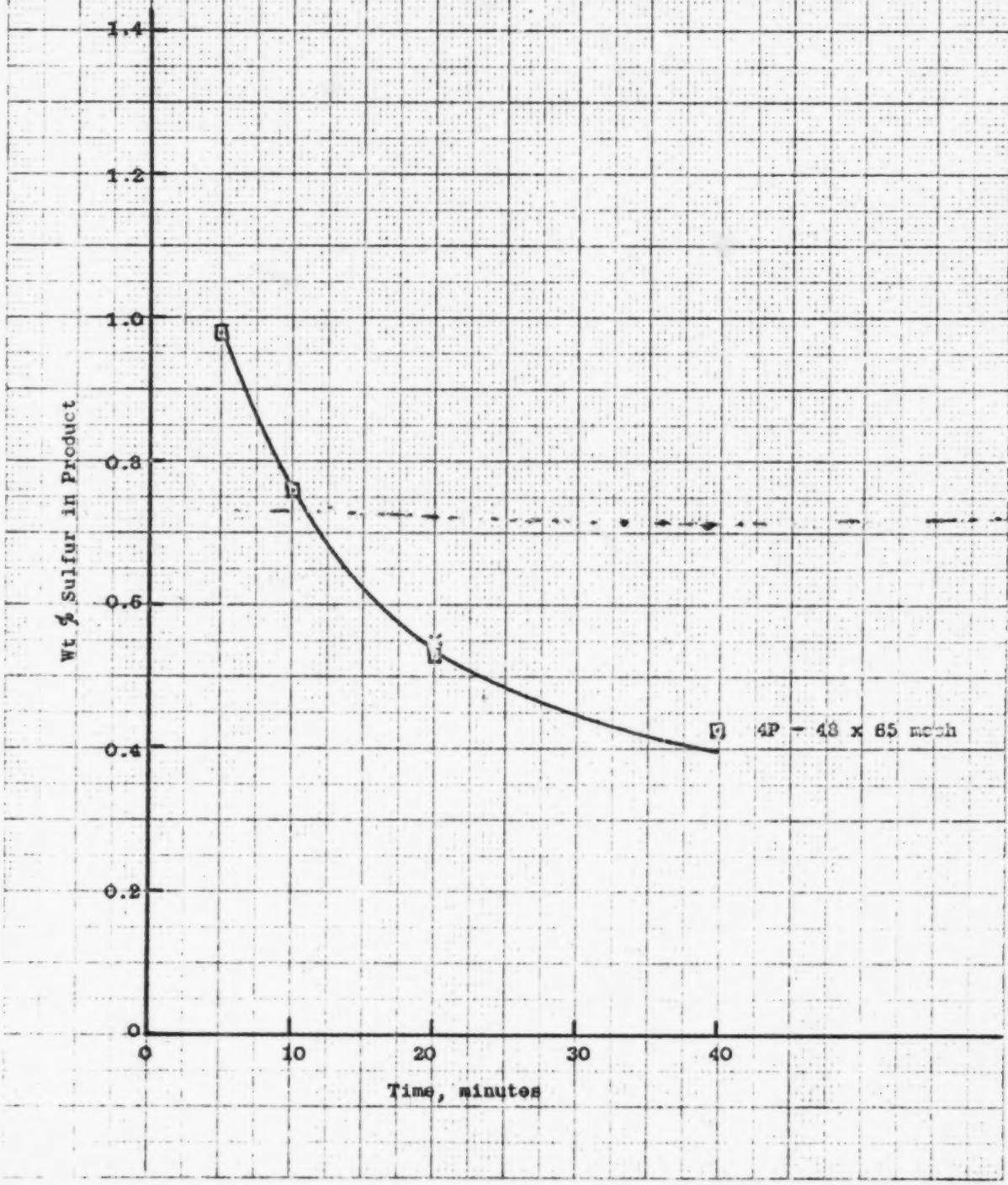
$$CaCO_3 + H_2S = CaS + H_2O + CO_2$$

becomes unfavorable. It would be necessary to use a metal oxide acceptor in this instance such as MnO as had been proposed previously.[5]

FIGURE 4

DIFFERENTIAL DESULFURIZATION OF PRECONDITIONED COAL AT 1500°F

Initial Sulfur Content = 3.49%



VI. ECONOMICS

The investment and operating costs for the cases considered here were not arrived at by a detailed estimation procedure but are only approximate. They were arrived at by ratioing unit costs as detailed in the Annual Report[1] to the equipment sizes required here.

A summary of the economics of the two cases considered here is given in comparison with the total gasification case, i.e., Case II of the Annual Report, in Table I of this report (cf. Summary). The figures are based on 1976 operation and include escalation in labor and materials, and interest during construction at the rate of 7-1/2% per year. The operating factor of the plant is taken at 70%. The product distributions for the cases are summarized in Table IV.

The investment break down by section, and manpower and utilities for two cases are given in Tables V-A and V-B. The power balances around the expander and air compressor are given in Table VI. Operating cost breakdowns are given in Tables VII-A and VII-B, respectively.

Figure 5 shows that the cost of the combined fuel product decreases progressively from 70¢/MM Btu to 60¢/MM Btu, with coal input at 40¢/MM Btu, as one goes from the total to minimum gasification case. The process thermal efficiency also increases in going through the same progression.

Where the total product is used as conventional boiler fuel, it does not appear that the process considered here is attractive relative to flue gas scrubbing. The estimated cost of sulfur removal for the minimum gasification case cited here is 20¢/MM Btu of fuel burned versus 15¢/MM Btu for the formate scrubbing case cited in the Annual Report.[1] It should also be noted, in making the comparison with flue gas scrubbing, that burning the char has an additional cost penalty associated with it of particulate removal. This must also be considered in comparing the merits of the total versus minimum gasification case. Total gasification lends itself to combustion processes which are both more efficient and cleaner.

The revised improved case cited earlier would effect some further cost improvements, however, the magnitude has not been estimated.

The major incentive for low-sulfur char co-product cases would be where a separate use existed for the low-sulfur char and where the gas would have a premium value. Such a case would comprise use of the char in a fluidized boiler and the gas in a combined cycle power plant.

Figure 5 has relevance to such a case and shows the cost of the low-sulfur char for the cases given in Table III as a function of the value of the gas. It is clear here that the minimum gasification case also produces the lowest cost char as long as the gas value is below about 75.5¢/MM Btu.

TABLE IV

Product Distribution

Case	This Study, Char as Co-Product: Minimum Gasification	This Study, Char as Co-Product: High Gasification	Annual Report to OAP: Case II Total Gasification
Coal Consumed			
Lb/hr	← 1,157,000 →		
MM Btu/hr (HHV)	← 13,812 →		
Products			
Gas			
Mols/hr	67,277	159,320	239,267
MM Btu/hr, HHV	5,598	8,273	10,415
Temperature, °F	462	512	665
Pressure, psia	← 25.7 →		
Btu/SCF, HHV	219	137	115
Char			
Lb/hr	604,300	365,500	0
MM Btu/hr, HHV	6,740	3,236	-
Sulfur Content, Wt. %	0.75	0.75	-
Sulfur, lb/hr	41,500	43,100	44,700

TABLE V-A

Investment, Manpower, and Utility Summary
Minimum Gasification Case

Section	Coal Preparation	Gasification	Sulfur Recovery-Solids Disposal	Totals
ISBL Investment, $ MM	5.0	20.3	13.9	39.2
Operators, Men/Shift	3	7	6	16
Utilities				
Electricity, KW	4,200	4,250(1)	15,520	23,970
Cooling Water, GPM	-	5,510	11,260	16,770
OSBL Investment, $ MM				
Offsites at 6.7% of ISBL Investment				2.6
Miscellaneous Utilities at 1% of ISBL Investment				0.4
Electrical				2.6
Cooling Water at 33 $/GPM				0.6
Sub-Total, OSBL				6.2
Installed Plant Cost, ISBL + OSBL, $ MM				45.4
			Escalation	8.5
				53.9
		Interest During Construction		9.0
			Total	62.9

(1) Does not include the main air compressors, which are driven directly by expanders.

TABLE V-B

Investment, Manpower, and Utility Summary
High Gasification Case

Section	Coal Preparation	Gasification	Sulfur Recovery-Solids Disposal	Totals
ISBL Investment, $ MM	5.0	48.6	14.2	67.8
Operators, Men/Shift	3	9	6	18
Utilities				
Electricity, KW	4,200	7,316(1)	15,420	26,936
Cooling Water, GPM	-	13,350	11,700	25,050
Boiler Feed Water, GPM	-	813	-	813
OSBL Investment, $ MM				
Offsites at 6.7% of ISBL Investment				4.5
Miscellaneous Utilities at 1% of ISBL Investment				0.7
Electrical				5.0
Cooling Water at 33 $/GPM				0.8
Boiler Feed Water at 380 $/GPM				0.3
Sub-Total, OSBL				11.3
Installed Plant Cost, ISBL + OSBL, $ MM				79.1
			Escalation	14.8
				93.9
			Interest During Construction	15.8
			Total	109.7

(1) Does not include main air compressors, which are driven directly by expanders.

TABLE VI

Expander Output, Power Output, and Air Compressor Requirement

Case	Minimum Gasification	Maximum Gasification
Output from Expanders, MM Btu/hr	286.0(1)	721.3(2)
Less 5% Mechanical Inefficiency, MM Btu/hr	-14.3	-36.1
	271.7	685.2
Less Main Air Compressor Requirement, MM Btu/hr	-153.8	-372.1
Net Power Generated, MM Btu/hr	117.9	313.1
Equivalent MW	34.5	91.7

(1) From Table A-VIII.

(2) From Table B-VII.

TABLE VII-A

Direct Operating Costs
Ex Coal Costs
Minimum Gasification Case

Direct Operating Costs (6,132 hr/yr)		$ MM/yr
1.	Direct Operating Labor 16 men/shift at (40,000 $/yr)/man/shift)	0.640
2.	Maintenance Labor 1.6% of Investment	0.726
	Sub-Total, Direct Labor	1.366
3.	Direct Supervision, 15% of 1 + 2	0.205
4.	Indirect Overhead, 50% of 1 + 2 + 3	0.786
5.	Payroll Overhead, 15% of 1 + 2 + 3 + 4	0.353
6.	Maintenance Materials, 2.4% of Investment	1.090
7.	Miscellaneous Supplies, 15% of Maintenance Materials	0.164
8.	Utilities, Chemicals and Catalyst	
	Electricity, 23,970 KW at 9 mills/KWH	1.512
	Cooling Water, 16,770 GPM at 2¢/1000 gal	0.123
	Make-up Acceptor, 6,600 lb/hr at 3.5 $/Ton	0.071
	K_2CO_3, 7.1 MM SCFH Gas Treated at (1,000 $/yr)/MM SCFH	0.006
	Total Direct Operating Costs	5.676
	Escalation	1.064
	Total	6.740

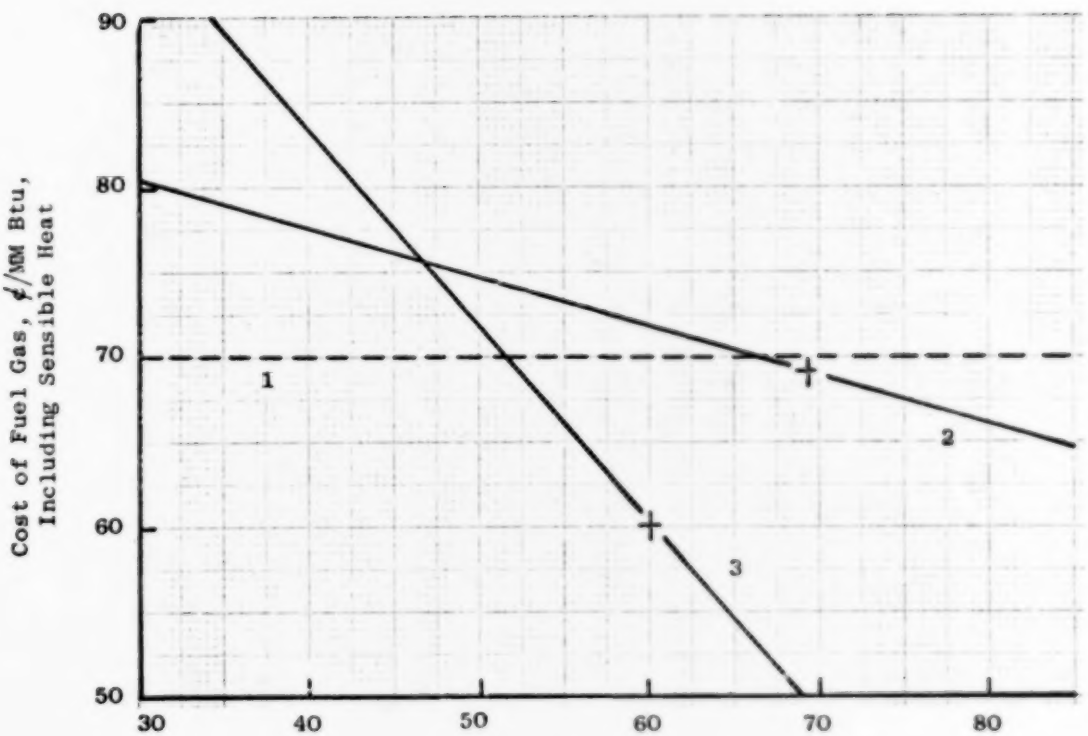
TABLE VII-B

Direct Operating Costs
Ex Coal Cost
High asification Case

Direct Operating Costs (6,132 hr/yr)	$ MM/yr
1. Direct Operating Labor 18 men/shift at (40,000 $/yr)/(man/shift)	0.720
2. Maintenance Labor 1.6% of Investment	1.085
Sub-Total, Direct Labor	1.805
3. Direct Supervision, 15% of 1 + 2	0.271
4. Indirect Overhead, 50% of 1 + 2 + 3	1.038
5. Payroll Overhead, 15% of 1 + 2 + 3 + 4	0.467
6. Maintenance Materials, 2.4% of Investment	1.627
7. Miscellaneous Supplies, 15% of Maintenance Materials	0.244
8. Utilities, Chemicals and Catalyst	
Electricity, 26,936 KW at 9 mills/KWH	1.699
Cooling Water, 25,050 GPM at 2¢/1000 gal	0.185
Boiler Feed Water, 813 GPM at 25¢/1000 gal	0.074
Make-up Acceptor, 6,800 lb/hr at 3.5 $/Ton	0.073
K_2CO_3, 6.7 MM SCFH Gas Treated at (1,000 $/yr)/MM SCFH	0.006
Total Direct Operating Costs	7.489
Escalation	1.404
Total	8.893

FIGURE 5

COST OF FUEL GAS AS A FUNCTION OF THE
PRICE RECEIVED FOR CHAR:COAL AT 40¢/MM BTU



1. Annual Report to OAP, Ref. 1 - Total gasification.
2. High gasification, char as a co-product.
3. Minimum gasification, char as a co-product.
   These points represent cost of combined fuel product to boiler.

It is noted in the cases reported here, as well as in Case II of the Annual Report,[1] that a major part of the total cost of the operation is in the sulfur recovery operation.

It is of interest to compare the costs given here, for the liquid-phase Claus system, with the more "conventional" once-through Claus system as costed in a recent report by Shell.[8]

To get the cost of the liquid-phase Claus system, one must exclude from the recovery system costs, the "Squires" reactors, the CO_2 recovery and handling system and the acceptor disposal systems uniquely associated with our system. Thus, the investment in the liquid-phase Claus system itself is only about 54% of the total recovery cost. The installed cost for the liquid-phase Claus unit to produce 450 long tons of sulfur/stream day is thus about $7.5 MM for the minimum gasification case or about $17 M/long ton of sulfur/stream day. The "comparable" Shell case[8] at 10 atm handling 2% H_2S gas in a "conventional" Claus has an installed cost, excluding utilities and offsites of $3.05 MM for 116 long tons of sulfur/stream day or about $26 M/long ton sulfur/stream day. The Shell system seems more expensive, but the bulk of the cost is tied up in the incinerator system which is not required in our particular case. The "conventional" Claus, as expected, is cheaper without the incinerator system, i.e., about $10 M/long ton of sulfur/stream day. The "conventional" Claus cannot be used in the "Squires" system because of the high steam content of the gas treated.

VII. BIBLIOGRAPHY

1. "Development of the CO_2 Acceptor Process Directed Towards Low-Sulfur Boiler Fuel." Curran, G.P.; Clancey, J.T.; Fink, C.E.; Pasek, Bedrich; Pell, Melvyn and Gorin, Everett. Annual Report from Consolidation Coal Co. to Control Systems Division, Office of Air Programs, Environmental Protection Agency, Under OAP Contract EHSD-71-15 for Period Sept. 1, 1970 to Nov. 1, 1971 NTIS Accession No. PB 210-840.

2. Consolidation Coal Co., Research & Development Report No. 16 to The Office of Coal Research, U.S. Dept. of the Interior Under Contract No. 14-01-0001-415, Interim Report No. 2, "Low-Sulfur Boiler Fuel Using the Consol CO_2 Acceptor Process." Nov. 1967. U.S. Dept. of Commerce, National Tech. Information Service PB-17610.

3. Ruth, L.A.; Squires, A.M.; and Graff, R.A.; "Desulfurization of Fuels With Calcined Dolomite: Part III - First Results for Reaction of $CaCO_3$ with H_2S." Paper presented at ACS Meeting, Los Angeles, March 31, 1971.

4. Zielke, C.W.; Curran, G.P.; Gorin, Everett; and Goring, G.E.; Ind. Eng. Chem., 46, 53 (1954).

5. Batchelor, J.D.; Gorin, Everett; and Zielke, C.W., Ind. Eng. Chem., 52, 161 (1960).

6. United States Patents Issued to Consolidation Coal Company:

   a. U.S. 2,717,868 - Sept. 13, 1955
   b. U.S. 2,824,047 - Feb. 18, 1958
   c. U.S. 2,927,063 - Aug. 23, 1960
   d. U.S. 2,950,229 - Aug. 23, 1960
   e. U.S. 2,950,231 - Aug. 23, 1960
   f. U.S. 3,101,303 - Aug. 20, 1963

7. Curran, G.P.; Fink, C.E.; and Gorin, Everett; "Production of Low-Sulfur Boiler Fuel - Application of CO_2 Acceptor Process." Paper presented before the Second International Conference on Fluidized Combustion, Hueston Woods, Ohio, Oct. 4-7, 1970.

8. Shpall, R.T., "Claus Technical and Economic Study." Shell Development Co. to Control Systems Division, Office of Air Programs, Environmental Protection Agency, Contract EHSD-71-15, April, 1972.

9. Consolidation Coal Company unpublished work.

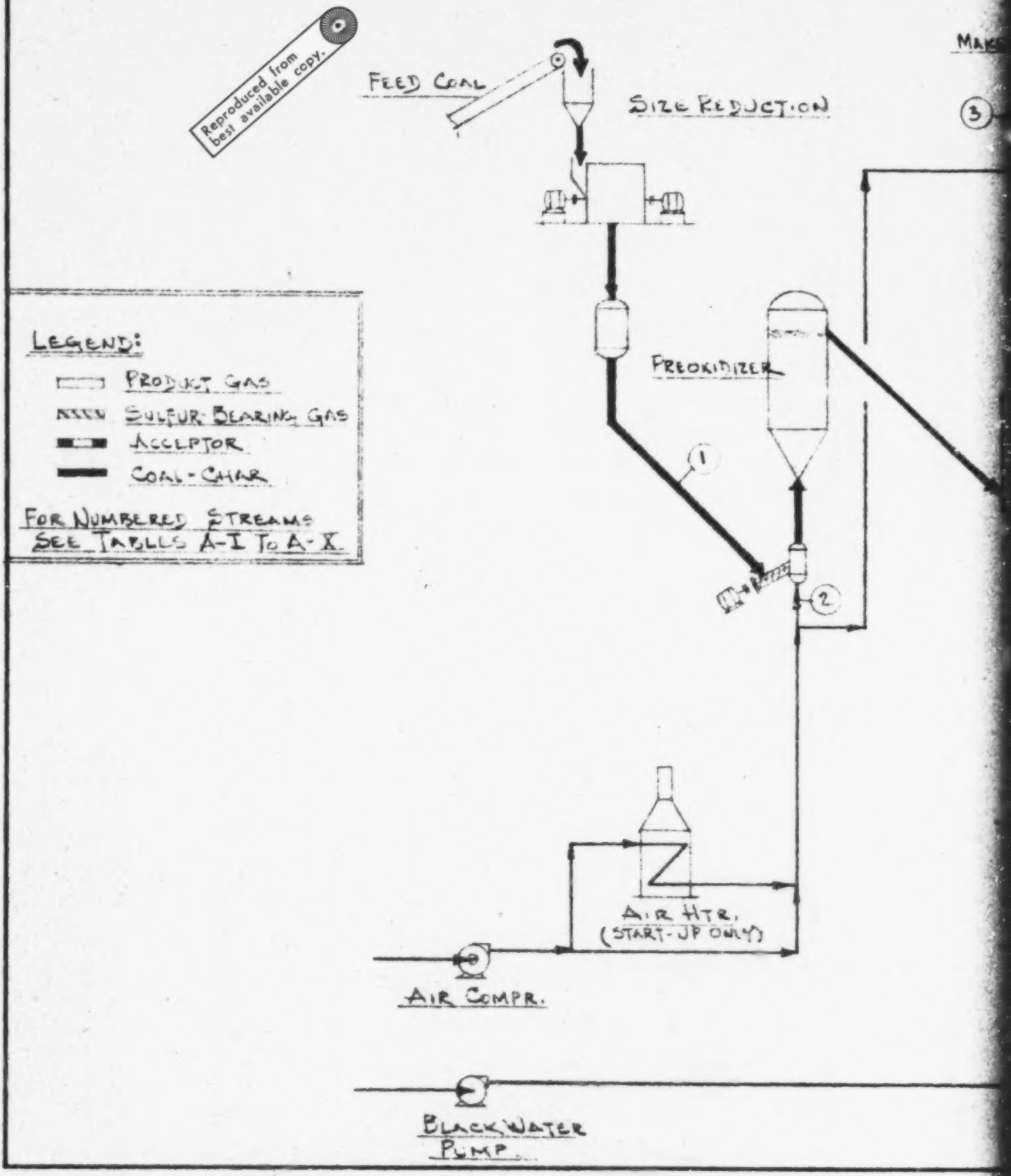
VIII. APPENDICES

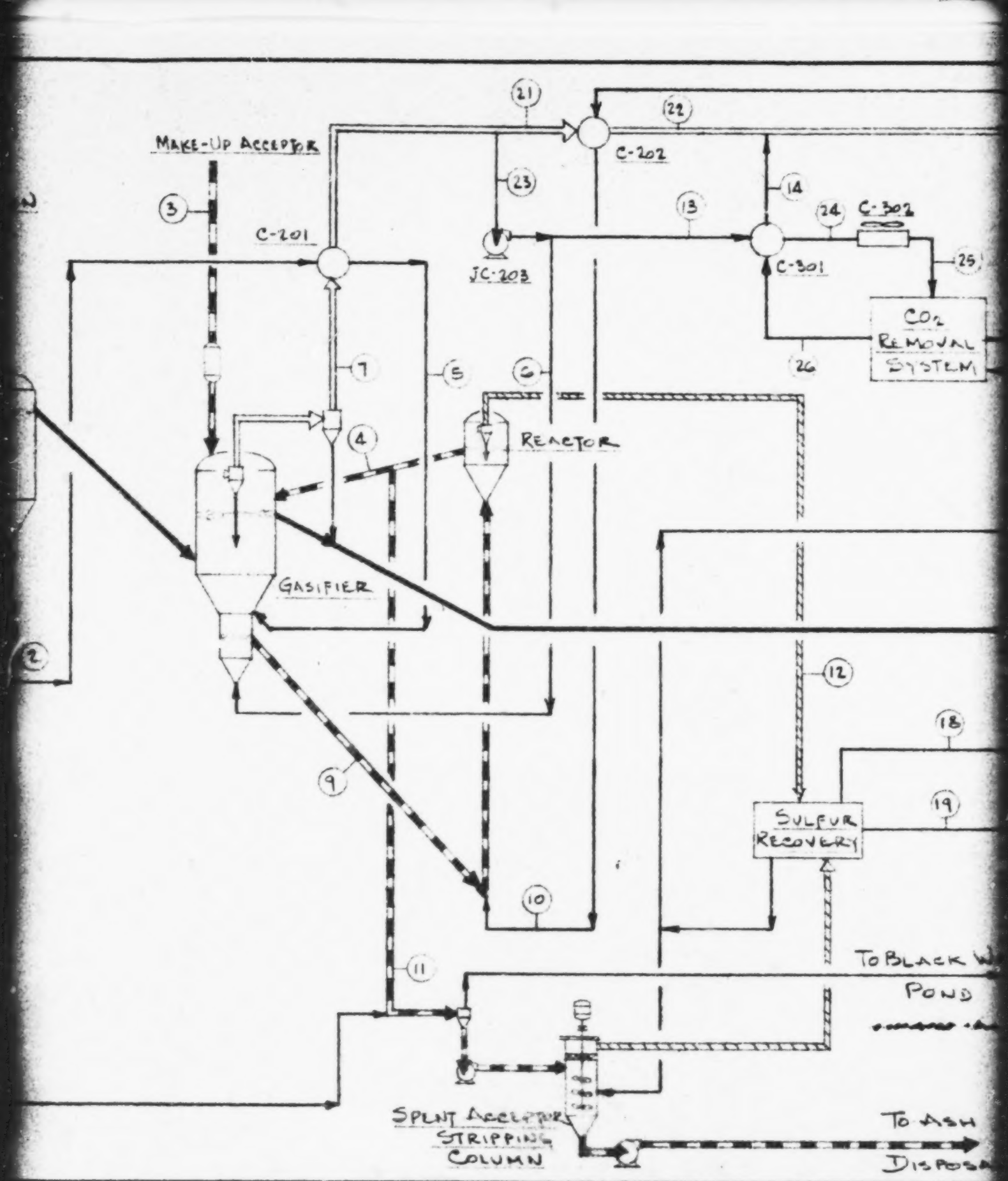
Appendix A

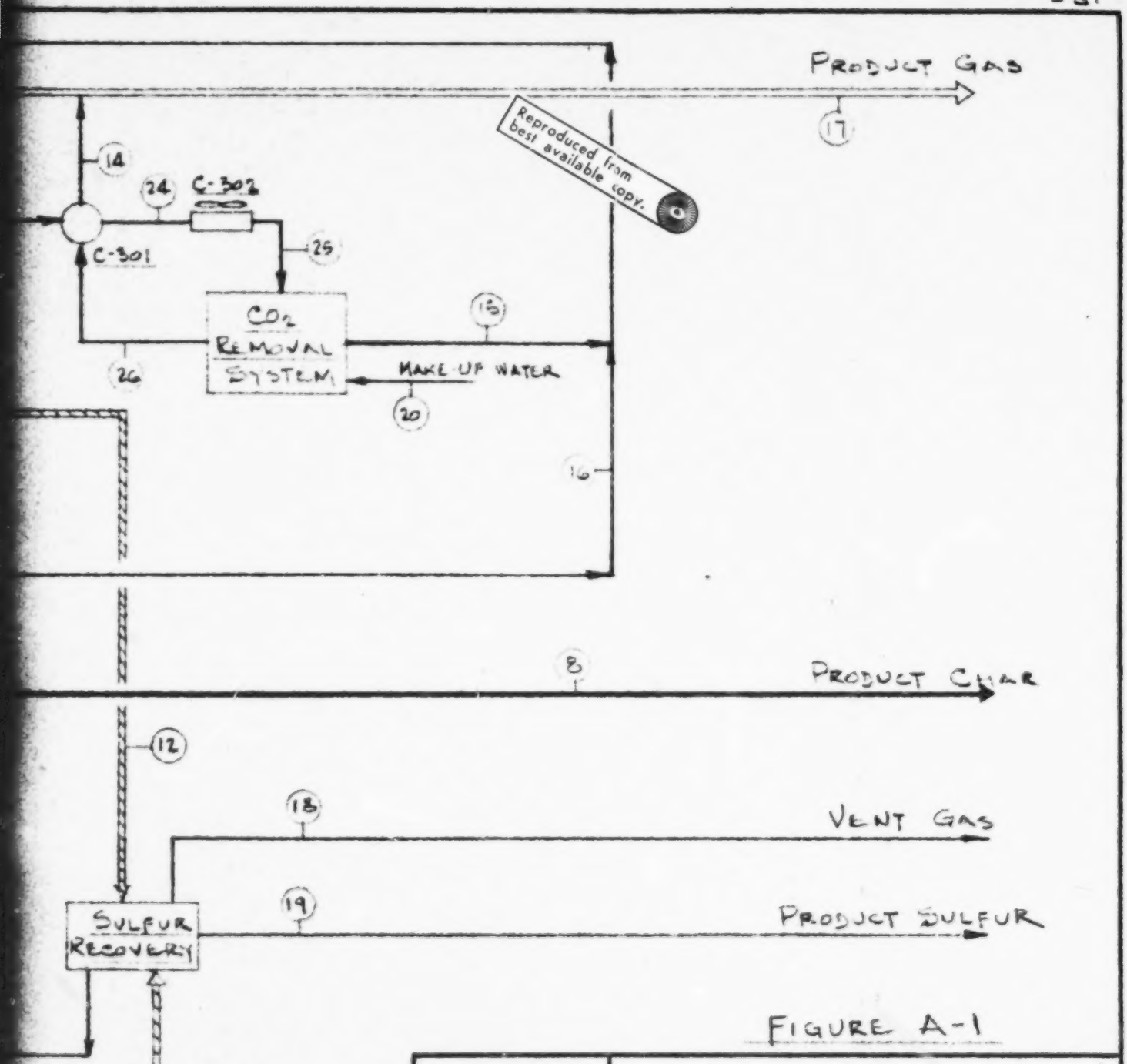
Mass and Heat Balances
Minimum Gasification Case

Figure A-1
Table A-I
A-II
A-III
A-IV
A-V
A-VI
A-VII
A-VIII
A-IX
A-X

Reproduced from best available copy.




MAKE-UP ACCEPTOR
C-201
C-202
C-301
C-302
JC-203
REACTOR
GASIFIER
CO_2 REMOVAL SYSTEM
SULFUR RECOVERY
SPENT ACCEPTOR STRIPPING COLUMN
TO ASH DISPOSAL
POND
1
2
3
4
5
6
7
9
10
11
12
13
14
18
19
21
22
23
24
25
26



Reproduced from best available copy.

FIGURE A-1

REVISIONS	CONSOLIDATION COAL CO. RESEARCH & DEVELOPMENT DIVISION LIBRARY, PENNA.		
	SCHEMATIC FLOW DIAGRAM		
	LOW SULFUR FUELS - CO_2 ACCEPTOR PROCESS		
	MINIMUM GASIFICATION CASE		
DRAWN 3/17/71	SCALE	PROJECT	DRAWING NO.
CHKD	NONE	553.00	XF-3255
APPR			

Table A-1

Mass and Heat Balance
Precxidizer - Gasifier
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Mol %	Elemental Balance, Pounds H	C	N	O	S	Ash or Inert	$MgO \cdot Ca$	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu	Heat (HHV) of Combustion MM Btu
Input															
Feed Coal	1														
MF Coal		1,087,600	-		52,200	759,200	13,000	82,600	46,800	133,800	-	60	0	0	13,812.4
Moisture		69,400	3,852		7,700	-	-	61,700	-	-	-	60	0	0	
Sub-Total		1,157,000			59,900	759,200	13,000	144,300	46,800	133,800	-		0	0	
Air to Precxidizer	2														
O_2		92,400	2,888	20.9	-	-	-	92,400	-	-	-	398	73.9	6.8	
N_2		304,500	10,869	78.7	-	-	304,500	-	-	-	-	398	84.4	25.7	
Moisture		1,100	60	0.4	100	-	-	1,000	-	-	-	398	1,213.5	1.3	
Sub-Total		398,000	13,817	100.0	100	-	304,500	93,400	-	-	-			33.8	
Air to Gasifier	5														
O_2		157,200	4,911	20.9	-	-	-	157,200	-	-	-	1,318	305.5	48.0	
N_2		517,900	18,482	78.7	-	-	517,900	-	-	-	-	1,318	327.6	169.7	
Moisture		1,800	101	0.4	200	-	-	1,600	-	-	-	1,318	1,672.7	3.0	
Sub-Total		676,900	23,494	100.0	200	-	517,900	158,800	-	-	-			220.7	
Makeup Acceptor	3														
$MgCO_3 \cdot CaCO_3$		6,100	33		-	800	-	2,600	-	-	2,700	60	0	0	
Inert		500	-		-	-	-	-	-	500	-	60	0	0	
Sub-Total		6,600			-	800	-	2,600	-	500	2,700		0	0	
Recycle Acceptor	4														
$MgO \cdot CaCO_3$		912,700	6,500		-	78,100	-	312,000	-	-	522,600	1,300	326.0	297.5	
Inert		110,600	-		-	-	-	-	-	110,600	-	1,300	300.0	33.2	
Sub-Total		1,023,300			-	78,100	-	312,000	-	110,600	522,600			330.7	
Recycle Gas	6	747,400	30,477		25,600	143,300	372,300	206,000	200	-	-	1,350	488.2	364.9	
Heat of Reaction:		HHV of Coal - (HHV of Net Gas + HHV of Char) = (13,812.4 - (5,595.3 + 6,740.0)) MM Btu =												1,477.1	
Totals		4,009,200	-	-	85,800	981,400	1,207,700	917,100	47,000	244,900	525,300			2,427.2	
Output															
Gas	7														
Net Gas															
CH_4		129,200	8,055	11.9	32,500	96,700	-	-	-	-	-	1,500	1,190	153.7	3,082.7
H_2		15,200	7,539	11.1	15,200	-	-	-	-	-	-	1,500	5,063	77.0	928.7
CO		344,200	12,287	18.1	-	147,600	-	196,600	-	-	-	1,500	382.8	131.8	1,495.5
CO_2		274,800	6,243	9.2	-	75,000	-	199,800	-	-	-	1,500	377.6	103.8	-
N_2		822,300	29,347	43.2			822,300	-				1,500	378.7	311.4	-
NH_3		8,700	511	0.7	1,500	-	7,200	-	-	-	-	1,500	994	8.6	84.1
H_2S		600	19	-	-	-	-	-	600			1,500	409.2	0.2	4.3
H_2O (v)		70,400	3,906	5.8	7,900	-	-	62,500	-	-	-	1,500	1,772.8	124.8	-
Sub-Total (Net Gas)		1,665,400	67,907	100.0	57,100	319,300	829,500	458,900	600	-	-			911.3	5,595.3
Recycle Gas		747,400	30,477	-	25,600	143,300	372,300	206,000	200	-	-	1,500		409.0	
Sub-Total (Stream 7)		2,412,800	98,384	-	82,700	462,600	1,201,800	664,900	800	-	-			1,320.3	
Char	8	604,300	-	-	3,100	456,000	5,900	1,000	1,500	133,800	-	1,500	499	301.5	6,740.0
Acceptor	9														
$MgO \cdot CaCO_3$		734,800	5,233	-	-	62,800	-	251,200	-	-	420,800	1,425	364.5	267.8	
$MgO \cdot CaS$		146,200	1,300	-	-	-	-	-	41,700	-	104,500	1,425	275.7	40.3	
Inert		111,100	-	-	-	-	-	-	-	111,100	-	1,425	334.4	37.1	
Sub-Total		992,100	6,533	-	-	62,800	-	251,200	41,700	111,100	525,300			345.2	
Heat of Reaction:		HHV of H_2S which reacted with Acceptor: 1,300 mol x 242,100 Btu/mol =												314.7	
		$H_2S + MgO \cdot CaCO_3 \longrightarrow MgO \cdot CaS + CO_2 + H_2O$ (ℓ): 1,300 mol x 30,620 Btu/mol =												39.8	
		Calcination of Makeup Acceptor, $MgCO_3 \cdot CaCO_3 \longrightarrow MgO \cdot CaCO_3 + CO_2$: 33 mol x 51,200 Btu/mol =												1.7	
Heat Loss														104.0	
Totals		4,009,200	-	-	85,800	981,400	1,207,700	917,100	47,000	244,900	525,300			2,427.2	

Table A-II

Mass and Heat Balance
Reactor (Squires)
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Acceptor	9					
$MgO \cdot CaCO_3$		734,800	5,233	1425	364.5	267.8
$MgO \cdot CaS$		146,200	1,300	1425	275.7	40.3
Inert		111,100	-	1425	334.4	37.1
Sub-Total		992,100	6,533			345.2
Gas	10					
CO_2		805,400	18,300	1147	273.6	220.4
H_2O (v)		329,700	18,300	1147	1,581.3	521.4
Sub-Total		1,135,100	36,600			741.8
Heat of Reaction						
$MgO \cdot CaS + CO_2 + H_2O$ (v) $\longrightarrow$ $MgO \cdot CaCO_3 + H_2S$						
1,300 mols x 30,620 Btu/mol =						39.8
Totals		2,127,200	-	-	-	1,126.8
Output						
Recycle Acceptor	4					
$MgO \cdot CaCO_3$		912,700	6,500	1300	326.0	297.5
Inert		110,600	-	1300	300.0	33.2
Sub-Total		1,023,300				330.7
Spent Acceptor	11					
$MgO \cdot CaCO_3$		4,600	33	1300	326.0	1.5
Inert		500	-	1300	300.0	0.1
Sub-Total		5,100				1.6
Gas	12					
CO_2		748,200	17,000	1300	318.1	238.0
H_2S		44,300	1,300	1300	341.7	15.1
H_2O (v)		306,300	17,000	1300	1,661.8	509.0
Sub-Total		1,098,800	35,300			762.1
Heat Loss						32.4
Totals		2,127,200	-	-		1,126.8

Table A-III

Mass and Heat Balance
Heat Exchanger C-201
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	7	2,412,800	98,384	1500	547.2	1,320.3
Output						
Gas	21					
CH_4		93,700	5,842	1327	1,003	94.0
H_2		11,000	5,468	1327	4,439	48.8
CO		249,600	8,912	1327	333.4	83.2
CO_2		199,300	4,528	1327	326.1	65.0
N_2		596,500	21,285	1327	330.1	196.9
NH_3		6,300	371	1327	852.3	5.4
H_2S		500	14	1327	353.0	0.2
H_2O		51,000	2,833	1327	1,677.5	85.6
Sub-Total		1,207,900	49,253			579.1
Gas	23					
CH_4		93,500	5,828	1327	1,003	93.8
H_2		11,000	5,454	1327	4,439	48.8
CO		249,000	8,890	1327	333.4	83.0
CO_2		198,700	4,517	1327	326.1	64.8
N_2		595,000	21,232	1327	330.1	196.4
NH_3		6,300	370	1327	852.3	5.4
H_2S		500	14	1327	353.0	0.2
H_2O		50,900	2,826	1327	1,677.5	85.4
Sub-Total		1,204,900	49,131			577.8
Heat Exchange						163.2
Heat Loss (1)						0.2
Totals		2,412,800	98,384			1,320.3
Low Temperature Side:						
Input						
Air	-					
O_2		157,200	4,911	398	73.9	11.6
N_2		517,900	18,482	398	84.4	43.7
H_2O (v)		1,800	101	398	1,213.5	2.2
Sub-Total		676,900	23,494			57.5
Heat Exchange						163.2
Totals						220.7
Output						
Air	5					
O_2		157,200	4,911	1318	305.5	48.0
N_2		517,900	18,482	1318	327.6	169.7
H_2O (v)		1,800	101	1318	1,672.7	3.0
Totals		676,900	23,494			220.7

(1) By difference, to force the heat balance.

Table A-IV

Mass and Heat Balance
Heat Exchanger C-202
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	21	1,207,900	49,253	1327	-	579.1
Output						
Gas	22					
CH_4		93,700	5,842	904	597.4	56.0
H_2		11,000	5,468	904	2,936	32.3
CO		249,600	8,912	904	216.5	54.0
CO_2		199,300	4,528	904	205.1	40.9
N_2		596,500	21,285	904	215.1	128.3
NH_3		6,300	371	904	531.9	3.4
H_2S		500	14	904	223.3	0.1
H_2O (v)		51,000	2,833	904	1,456.2	74.3
Sub-Total		1,207,900	49,253			389.3
Heat Exchange						189.5
Heat Loss(1)						0.3
Totals		1,207,900	49,253			579.1
Low Temperature Side:						
Input						
Gas	15					
CO_2		57,200	1,300	200	29.2	1.7
H_2O (v)		700	40	200	1,124.9	0.8
Sub-Total		57,900	1,340			2.5
Gas	16					
CO_2		748,200	17,000	676	143.9	107.7
H_2O (v)		329,000	18,260	676	1,343.9	442.1
Sub-Total		1,077,200	35,260			549.8
Heat Exchange						189.5
Totals		1,135,100	36,600			741.8
Output						
Gas	10					
CO_2		805,400	18,300	1147	273.6	220.4
H_2O (v)		329,700	18,300	1147	1,581.3	521.4
Totals		1,135,100	36,600			741.8

(1) By difference, to force the heat balance.

Table A-V

Mass and Heat Balance
Compressor JC-203
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Gas	23	1,204,900	49,131	1327	-	577.8
Isentropic Work = $\bar{c}_p T_1 \left[(P_2/P_1)^{(N-1)/N} - 1\right]$ (49,131 mols) =						9.4(1)
Inefficiency = (9.4/0.89) - 9.4 =						1.2(2)
Totals		1,204,900	49,131	-	-	588.4
Output						
Recycle Gas	6					
CH_4		38,000	3,615	1350	1,027	59.6
H_2		6,800	3,383	1350	4,521	30.7
CO		154,500	5,515	1350	339.9	52.5
CO_2		123,200	2,802	1350	332.8	41.0
N_2		369,100	13,170	1350	336.5	124.2
NH_3		3,900	230	1350	870.8	3.4
H_2S		300	9	1350	360.3	0.1
H_2O		31,600	1,753	1350	1,690.1	53.4
Sub-Total		747,400	30,477			364.9
Gas to CO_2 Plant	13					
CH_4		35,500	2,213	1350	1,027	36.5
H_2		4,200	2,071	1350	4,521	19.0
CO		94,500	3,375	1350	339.9	32.1
CO_2		75,500	1,715	1350	332.8	25.1
N_2		225,900	8,062	1350	336.5	76.0
NH_3		2,400	140	1350	870.8	2.1
H_2S		200	5	1350	360.3	0.1
H_2O		19,300	1,073	1350	1,690.1	32.6
Sub-Total		457,500	18,654			223.5
Totals		1,204,900	49,131			588.4

(1) $\bar{c}_p$ = 9.318 Btu/lb mol °R
T_1 = 1787°R
N = $\bar{c}_p/(\bar{c}_p - 1.99)$ = 1.2731
P_2/P_1 = 1.0550 ($\Delta P \sim 12$ psi)

(2) 89% efficiency.

Table A-VI

Mass and Heat Balance
Heat Exchanger C-301
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas to CO_2 Plant	13	457,500	18,654	1350	-	223.5
Output						
Gas to CO2 Plant	24					
CH_4		35,500	2,213	524	293.5	10.4
H_2		4,200	2,071	524	1,605	6.7
CO		94,500	3,375	524	116.6	11.0
CO_2		75,500	1,715	524	105.1	7.9
N_2		225,900	8,062	524	116.3	26.3
NH_3		2,400	140	524	275.5	0.7
H_2S		200	5	524	117.0	-
H_2O		19,300	1,073	524	1,271.7	24.5
Sub-Total		457,500	18,654			87.5
Heat Exchange						135.9
Heat Loss(1)						0.1
Totals		457,500	18,654			223.5
Low Temperature Side:						
Input						
Gas from CO2 Plant	26					
CH_4		35,500	2,213	230	97.5	3.5
H_2		4,200	2,071	230	585.1	2.5
CO		94,500	3,375	230	42.3	4.0
CO_2		18,300	415	230	35.8	0.7
N_2		225,900	8,062	230	42.3	9.6
NH_3		2,400	140	230	97.3	0.2
H_2S		200	5	230	41.3	-
H_2O		31,400	1,743	230	1,138.1	35.7
Sub-Total		412,400	18,024			56.2
Heat Exchange						135.9
Totals		412,400	18,024			192.1
Output						
Gas from CO_2 Plant	14					
CH_4		35,500	2,213	1150	824.8	29.3
H_2		4,200	2,071	1150	3,806	16.0
CO		94,500	3,375	1150	283.8	26.8
CO_2		18,300	415	1150	274.5	5.0
N_2		225,900	8,062	1150	281.2	63.5
NH_3		2,400	140	1150	713.7	1.7
H_2S		200	5	1150	297.3	0.1
H_2O		31,400	1,743	1150	1,582.9	49.7
Totals		412,400	18,024			192.1

(1) By difference, to force the heat balance.

Table A-VIII

Mass and Heat Balance for Expander
Product Gas at 25.7 psia
Minimum Gasification Case; See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Gas	22	1,207,900	49,253	904	-	389.3
Gas from CO_2 Plant	14	412,400	18,024	1150	-	192.1
Totals		1,620,300	67,277			581.4
Output						
Product Gas	17					
CH_4		129,200	8,055	969	655.1	84.6
H_2		15,200	7,539	969	3,165	48.1
CO		344,100	12,287	969	234.1	80.6
CO_2		217,600	4,943	969	223.2	48.6
N_2		822,400	29,347	969	232.3	191.0
NH_3		8,700	511	969	578.7	5.0
H_2S		700	19	969	242.4	0.2
H_2O (v)		82,400	4,576	969	1,489.2	122.7
Sub-Total		1,620,300	67,277			580.8
Heat Loss(1)						0.6
Totals		1,620,300	67,277			581.4

(1) By difference, to force the heat balance.

Table A-VIII

Mass and Heat Balance for Expander
Product Gas at 25.7 psia
Minimum Gasification Case: See Figure A-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Product Gas at 206 psia	17	1,620,300	67,277	969	-	580.8
Output						
Work = $-0.91\ \bar{c}_p\ T_1 \left[(P_2/P_1)^{(N-1)/N} - 1\right]$ (67,277 mols) =						286.0(1)
Product Gas at 25.7 psia						
CH_4		129,200	8,055	462	249.1	32.2
H_2		15,200	7,539	462	1,389	21.1
CO		344,100	12,287	462	100.8	34.7
CO_2		217,600	4,943	462	89.8	19.5
N_2		822,400	29,347	462	100.5	82.7
NH_3		8,700	511	462	236.3	2.1
H_2S		700	19	462	100.0	0.1
H_2O		82,400	4,576	462	1,242.9	102.4
Sub-Total		1,620,300	67,277			294.8
Totals		1,620,300	67,277			580.8

(1) 91% efficiency
$\bar{c}_p$ = 8.385 Btu/lb mol °R
T_1 = 1429 °R
$N = \bar{c}_p/(\bar{c}_p - 1.99) = 1.3112$; $(N-1)/N = 0.2373$
P_2/P_1 = 25.7/206 = 0.1248

Table A-IX

Mass Balance
CO_2 Removal System
Minimum Gasification Case: See Figure A-1

Basis: 1 hr.

	Stream	Pounds	Mols	Temp. °F
Input				
Gas to CO_2 Plant	25			230
CH_4		35,500	2,213	
H_2		4,200	2,071	
CO		94,500	3,375	
CO_2		75,500	1,715	
N_2		225,900	8,062	
NH_3		2,400	140	
H_2S		200	5	
H_2O (v)		19,300	1,073	
Sub-Total		457,500	18,654	
Make-up Water	20	12,800	710	
Totals		470,300	19,364	
Output				
Gas from CO_2 Plant	26			230
CH_4		35,500	2,213	
H_2		4,200	2,071	
CO		94,500	3,375	
CO_2		18,300	415	
N_2		225,900	8,062	
NH_3		2,400	140	
H_2S		200	5	
H_2O (v)		31,400	1,743	
Sub-Total		412,400	18,024	
Make-up CO_2	15			200
CO_2		57,200	1,300	
H_2O (v)		700	40	
Sub-Total		57,900	1,340	
Totals		470,300	19,364	

Table A-X

Mass Balance
Sulfur Recovery
Minimum Gasification Case: See Figure A-1[1]

Basis: 1 hour

	Stream	Pounds	Mols	Temp. °F
Input				
Gas to Sulfur Recovery	12			1300
CO_2		748,200	17,000	
H_2S		44,300	1,300	
H_2O (v)		306,300	17,000	
Sub-Total		1,098,800	35,300	
Combustion Air	-			
O_2		21,000	657	
N_2		69,200	2,471	
H_2O		200	14	
Sub-Total		90,400	3,142	
Totals		1,189,200	-	
Output				
Gas	16			676
CO_2		748,200	17,000	
H_2O (v)		329,000	18,260	
Sub-Total		1,077,200	35,260	
Vent Gas	18			90
SO_2		400	7	
N_2		69,200	2,471	
H_2O (v)		100	8	
Sub-Total		69,700	2,486	
Product Sulfur	19	41,500	1,293	
Make H_2O	-	800	46	
Totals		1,189,200	-	

(1) Small streams to and from the Spent Acceptor Stripping Column have been neglected.

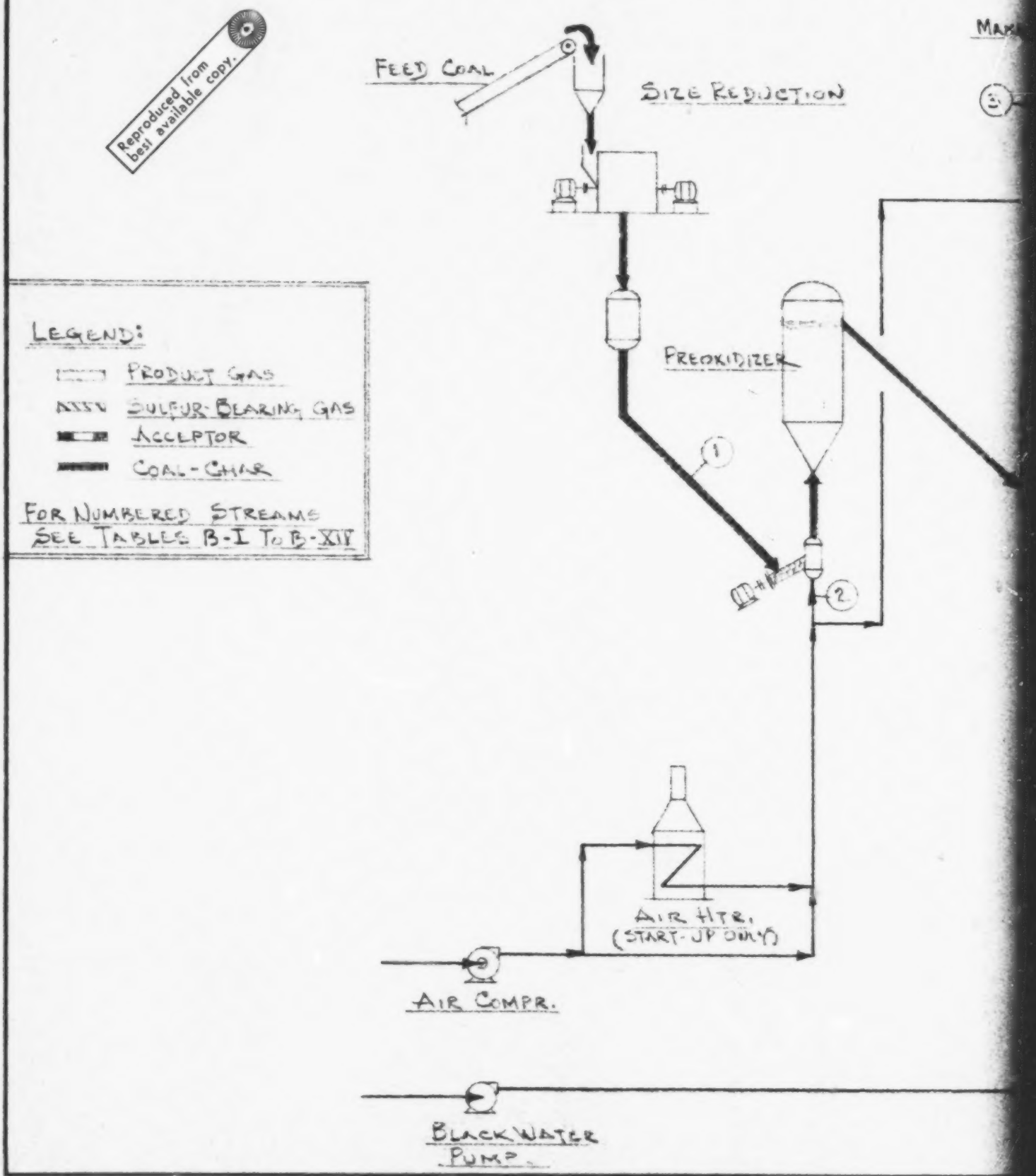
VIII. APPENDICES - Cont'd.

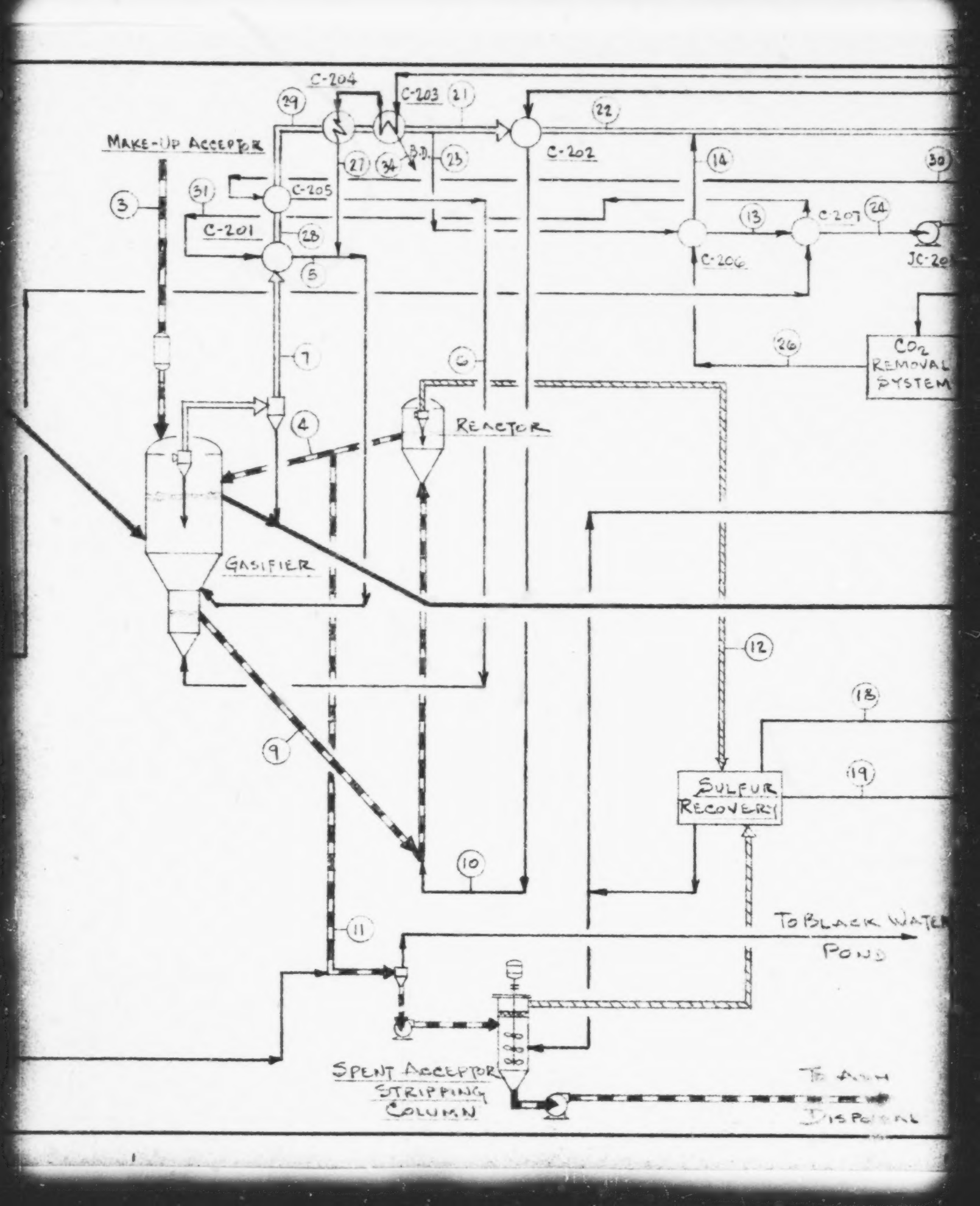
Appendix B

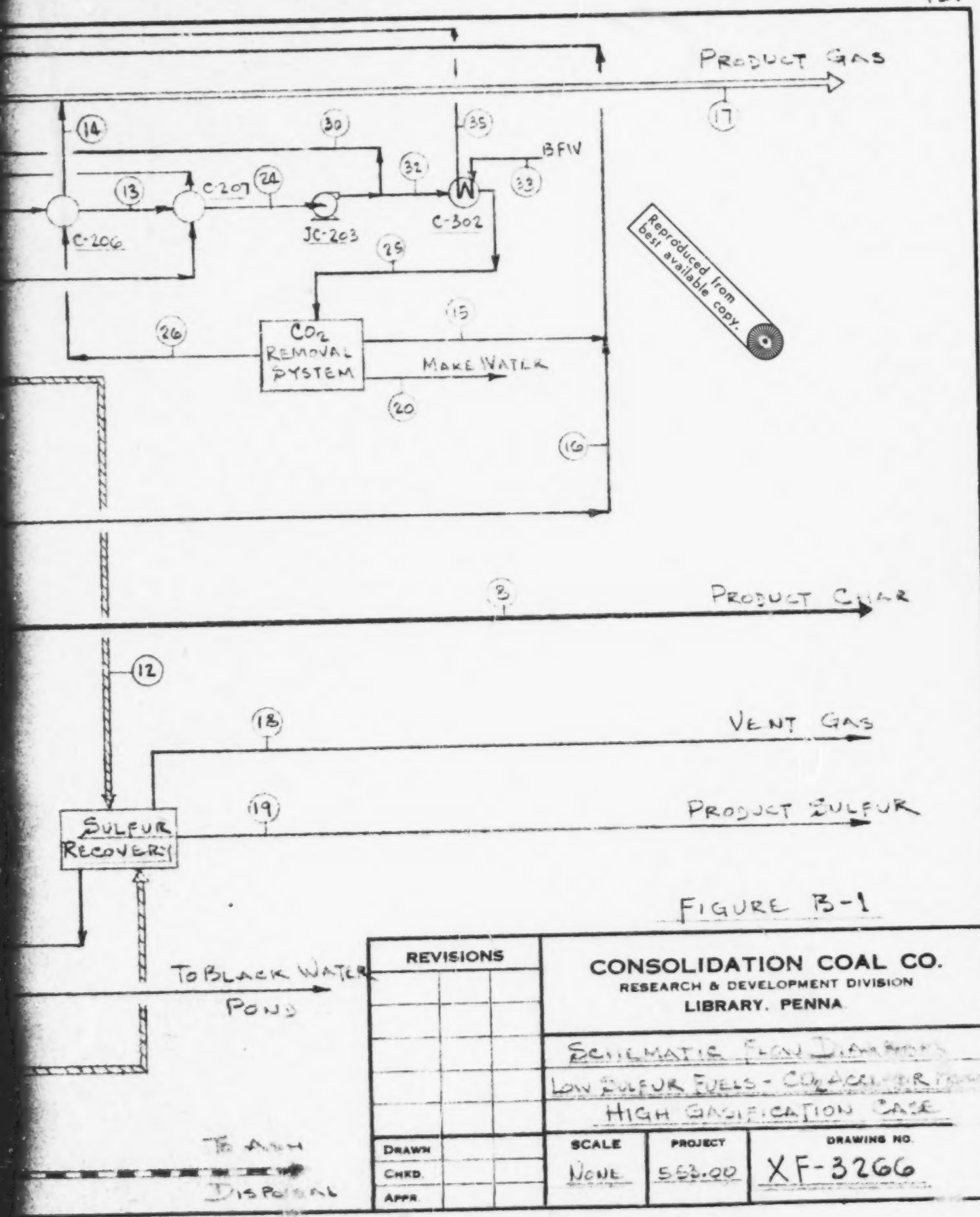
Mass and Heat Balances
High Gasification Case

Figure B-1
Table B-I
B-II
B-III
B-IV
B-V
B-VI
B-VII
B-VIII
B-IX
B-X
B-XI
B-XII
B-XIII
B-XIV

Reproduced from best available copy.




C-204
C-203
29
21
22
MAKE-UP ACCEPTOR
27
34
B.D.
23
C-202
14
30
31
C-205
3
C-201
28
13
C-207
24
C-206
JC-20
5
7
6
26
CO_2 REMOVAL SYSTEM
4
REACTOR
GASIFIER
12
18
9
19
SULFUR RECOVERY
10
TO BLACK WATER POND
11
SPENT ACCEPTOR STRIPPING COLUMN
TO ASH DISPOSAL



Reproduced from best available copy.

FIGURE B-1

REVISIONS		

CONSOLIDATION COAL CO.
RESEARCH & DEVELOPMENT DIVISION
LIBRARY, PENNA.

SCHEMATIC FLOW DIAGRAMS
LOW SULFUR FUELS - CO_2 ACCEPTOR PROCESS
HIGH GASIFICATION CASE

DRAWN	SCALE	PROJECT	DRAWING NO.
CHKD.	NONE	553.00	XF-3266
APPR.			

Table B-I

Mass and Heat Balance
Preoxidizer - Gasifier
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (4)

	Stream	Pounds	Mols	Mol %	Elemental Balance, Pounds H	C	N	O	S	Ash or Inert	$MgO \cdot Ca$	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu	Heat (HHV) of Combustion MM Btu
Input															
Feed Coal	1														
MF Coal		1,087,400	-	-	52,300	739,200	13,000	82,600	46,800	133,800	-	60	0	0	13,812.4
Moisture		66,400	3,688	-	7,700	-	-	61,700	-	-	-	60	0	0	
Sub-Total		1,153,800			59,900	739,200	13,000	144,300	46,800	133,800	-		0	0	
Air to Preoxidizer	2														
O_2		92,400	2,888	21.5		-	-	92,400	-	-	-	398	73.9	6.8	
N_2		304,500	10,869	78.8	-	-	304,500	-	-	-	-	398	84.4	25.7	
Moisture		1,000	59	0.4	100	-	-	900	-	-	-	398	1,213.5	1.2	
Sub-Total		397,900	13,816	100.0	100	-	304,500	93,300	-	-	-			33.7	
Air to Gasifier	5														
O_2		511,800	15,994	20.9	-	-	-	511,800	-	-	-	1,200	275.0	140.7	
N_2		1,686,200	60,186	78.7	-	-	1,686,200	-	-	-	-	1,200	294.7	496.9	
Moisture		6,000	329	0.4	700	-	-	5,300	-	-	-	1,200	1,609.3	9.7	
Sub-Total		2,204,000	76,509	100.0	700	-	1,686,200	517,100	-	-	-			647.3	
Steam to Gasifier	27	386,000	21,426	-	43,200	-	-	342,800	-	-	-	1,200	1,609.3	621.2	
Makeup Acceptor	3														
$MgCO_3 \cdot CaCO_3$		6,200	34	-	-	800	-	2,700	-	-	2,700	60	0	0	
Inert		600	-	-	-	-	-	-	-	600	-	60	0	0	
Sub-Total		6,800			-	800	-	2,700	-	600	2,700		0	0	
Recycle Acceptor	4														
$MgO \cdot CaCO_3$		947,400	6,747	-	-	81,000	-	323,900	-	-	542,500	1,300	326.0	308.9	
Inert		114,800	-	-	-	-	-	-	-	114,800	-	1,300	300.0	34.4	
Sub-Total		1,062,200			-	81,000	-	323,900	-	114,800	542,500			343.3	
Recycle Gas	6	1,714,900	72,283	-	46,100	249,300	897,200	521,900	400	-	-	1,200	458.9	787.4	
Heat of Reaction:		HHV of Coal - (HHV of Hot Gas + HHV of Char) = (13,812.4 - (8,272.8 + 3,236.4)) MM Btu =												2,303.2	
Totals		6,928,800	-	-	150,000	1,090,300	2,900,900	1,946,000	47,200	249,200	545,200			4,736.1	
Output															
Hot Gas	7														
CH_4		55,500	3,459	2.1	14,000	41,500	-	-	-	-	-	1,700	1,442	80.0	1,323.8
H_2		53,200	26,387	16.4	53,200	-	-	-	-	-	-	1,700	5,774	307.2	3,251.1
CO		830,400	29,644	18.4	-	356,100	-	474,300	-	-	-	1,700	440.8	366.0	3,607.8
CO_2		377,500	13,123	8.2	-	157,600	-	419,900	-	-	-	1,700	438.3	253.1	-
N_2		1,890,600	71,033	44.2	-	-	1,890,600	-	-	-	-	1,700	435.9	867.7	-
NH_3		8,700	511	0.3	1,500	-	7,200	-	-	-	-	1,700	1,137	9.8	69.9
H_2S		900	26	-	100	-	-	-	800	-	-	1,700	472.2	0.4	6.2
H_2O (v)		301,800	16,751	10.4	33,800	-	-	268,000	-	-	-	1,700	1,884.9	568.9	-
Sub-Total (Hot Gas)		3,818,600	160,936	100.0	102,600	555,200	1,897,800	1,162,200	800	-	-			2,453.1	8,272.8
Recycle Gas		1,714,900	72,283	-	46,100	249,300	897,200	521,900	400	-	-	1,700		1,101.7	
Sub-Total (Stream 7)		5,533,500	233,219	-	148,700	804,500	2,795,000	1,684,100	1,200	-	-			3,554.8	
Char	8	369,300	-	-	1,300	220,600	5,900	1,200	2,700	133,800	-	1,700	845.8	159.5	3,236.4
Acceptor	9														
$MgO \cdot CaCO_3$		782,600	5,431		-	65,200	-	260,700	-	-	436,700	1,700	452.0	344.7	
$MgO \cdot CaS$		151,800	1,358		-	-	-	-	43,300	-	108,500	1,700	335.3	50.9	
Inert		115,400	-		-	-	-	-	-	115,400	-	1,700	411.6	47.5	
Sub-Total		1,049,800			-	65,200	-	260,700	43,300	115,400	545,200			443.1	
Heat of Reaction:		HHV of H_2S which reacted with Acceptor: 1,350 mol x 242,100 Btu/mol =												326.8	
		$H_2S + MgO \cdot CaCO_3 \rightarrow MgO \cdot CaS + CO_2 + H_2O$ (4): 1,350 mol x 30,600 Btu/mol =												41.3	
		Calcination of Makeup Acceptor, $MgCO_3 \cdot CaCO_3 \rightarrow MgO \cdot CaCO_3 + CO_2$: 34 mol x 51,200 Btu/mol =												1.7	
Heat Loss														168.9	
Totals		6,928,800	-	-	150,000	1,090,300	2,900,900	1,946,000	47,200	249,200	545,200			4,736.1	

Table B-II

Mass and Heat Balance
Reactor (Squires)
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Acceptor	9					
$MgO \cdot CaCO_3$		762,600	5,431	1700	452.0	344.7
$MgO \cdot CaS$		151,800	1,350	1700	335.3	50.9
Inert		115,400	-	1700	411.6	47.5
Sub-Total		1,029,800				443.1
Gas	10					
CO_2		836,600	19,009	943	215.9	180.6
H_2O (v)		342,500	19,009	943	1,476.0	505.5
Sub-Total		1,179,100				686.1
Heat of Reaction						
$MgO \cdot CaS + CO_2 + H_2O(v) \rightarrow MgO \cdot CaCO_3 + H_2S$						
1,350 mols x 30,620 Btu/mol =						41.3
Totals		2,208,900				1,170.5
Output						
Recycle Acceptor	4					
$MgO \cdot CaCO_3$		947,400	6,747	1300	326.0	308.9
Inert		114,800	-	1300	300.0	34.4
Sub-Total		1,062,200				343.3
Spent Acceptor	11					
$MgO \cdot CaCO_3$		4,700	34	1300	326.0	1.5
Inert		600	-	1300	300.0	0.2
Sub-Total		5,300				1.7
Gas	12					
CO_2		777,200	17,659	1300	318.1	247.2
H_2S		46,000	1,350	1300	341.7	15.7
H_2O (v)		318,200	17,659	1300	1,661.8	528.8
Sub-Total		1,141,400	36,668			791.7
Heat Loss						33.8
Totals		2,208,900				1,170.5

Table B-III

Mass and Heat Balance
Heat Exchanger C-201
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	7	5,533,500	233,239	1700	-	3,554.8
Output	28					
CH_4		80,400	5,012	1627	1,334	107.3
H_2		77,100	38,237	1627	5,526	426.1
CO		1,203,300	42,960	1627	419.5	504.8
CO_2		836,900	19,016	1627	415.9	348.1
N_2		2,884,600	102,962	1627	414.9	1,196.8
NH_3		12,600	740	1627	1,102	13.9
H_2S		1,300	38	1627	451.5	0.6
H_2O		437,300	24,274	1627	1,844.6	806.6
Sub-Total		5,533,500	233,239			3,404.2
Heat Exchange						149.6
Heat Loss(1)						1.0
Totals		5,533,500	233,239			3,554.8
Low Temperature Side:						
Input						
Air	31					
O_2		511,800	15,994	946	209.8	107.4
N_2		1,686,200	60,186	946	226.2	381.4
H_2O (v)		6,000	329	946	1,477.5	8.9
Sub-Total		2,204,000	76,509			497.7
Heat Exchange						149.6
Totals		2,204,000	76,509			647.3
Output						
Air	5					
O_2		511,800	15,994	1200	275.0	140.7
N_2		1,686,200	60,186	1200	294.7	496.9
H_2O (v)		6,000	329	1200	1,609.3	9.7
Totals		2,304,000	76,509			647.3

(1) By difference, to force the heat balance.

Table B-IV

Mass and Heat Balance
Heat Exchanger C-205
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	28	5,533,500	223,239	1627	-	3,404.2
Output						
Gas	29					
CH_4		80,400	5,012	1483	1,171	94.1
H_2		77,100	38,237	1483	5,001	385.6
CO		1,203,300	42,960	1483	377.9	454.7
CO_2		836,900	19,016	1483	372.4	311.7
N_2		2,884,600	102,962	1483	373.9	1,078.6
NH_3		12,600	740	1483	979.7	12.3
H_2S		1,300	38	1483	403.6	0.5
H_2O		437,300	24,274	1483	1,763.4	771.1
Sub-Total		5,533,500	233,239			3,108.6
Heat Exchange						294.6
Heat Loss(1)						1.0
Totals		5,533,500	233,239			3,404.2
Low Temperature Side:						
Input						
Recycle Gas	30	1,714,900	72,283	700	-	492.8
Heat Exchange						294.6
Totals		1,714,900	72,283		-	787.4
Output						
Recycle Gas	6					
CH_4		24,900	1,553	1200	873.9	21.8
H_2		23,900	11,850	1200	3,983	95.2
CO		372,900	13,314	1200	297.6	111.0
CO_2		259,400	5,893	1200	289.0	74.9
N_2		894,000	31,909	1200	294.7	263.4
NH_3		3,900	229	1200	752.2	2.9
H_2S		400	12	1200	312.7	0.1
H_2O		135,500	7,523	1200	1,609.3	218.1
Totals		1,714,900	72,283			787.4

(1) By difference, to force the heat balance.

Table B-V

Mass and Heat Balance
Heat Exchangers C-203 and C-204
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	29	5,533,500	233,239	1483	-	3,108.6
Output						
Gas to C-202	21					
CH_4		49,400	3,082	1200	873.9	43.2
H_2		47,400	23,512	1200	3,983	188.8
CO		739,900	26,416	1200	297.6	220.2
CO_2		514,600	11,693	1200	289.0	148.7
N_2		1,773,700	63,310	1200	294.7	522.7
NH_3		7,700	455	1200	752.2	5.8
H_2S		800	23	1200	312.7	0.3
H_2O		268,900	14,926	1200	1,609.3	432.8
Sub-Total		3,402,400	143,417			1,562.5
Gas to C-206	23					
CH_4		31,000	1,930	1200	873.9	27.1
H_2		29,700	14,725	1200	3,983	118.3
CO		463,400	16,544	1200	297.6	137.9
CO_2		322,300	7,323	1200	289.0	93.1
N_2		1,110,900	39,652	1200	294.7	327.4
NH_3		4,900	285	1200	752.2	3.7
H_2S		500	15	1200	312.7	0.2
H_2O		168,400	9,348	1200	1,609.3	271.0
Sub-Total		2,131,100	89,822			978.7
Heat Exchange						566.0
Heat Loss(1)						1.4
Totals		5,533,500	233,239			3,108.6
Low Temperature Side:						
Input						
Boiler Feed Water	35	406,300	22,554	214	154	62.1
Heat Exchange						566.0
Totals		406,300	22,554			628.1
Output						
Blowdown	34	20,300	1,128	394	340.2	6.9
Steam	27	386,000	21,426	1200	1,609.3	621.2
Totals		406,300	22,554			628.1

(1) By difference, to force the heat balance.

Table B-VI

Mass and Heat Balance
Heat Exchanger C-202
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	21	3,402,400	143,417	1200	-	1,562.5
Output						
Gas	22					
CH_4		49,400	3,082	1104	780.3	38.5
H_2		47,400	23,512	1104	3,643	172.7
CO		739,900	26,416	1104	271.0	200.5
CO_2		514,600	11,693	1104	261.3	134.5
N_2		1,773,700	63,310	1104	268.7	476.6
NH_3		7,700	455	1104	678.8	5.2
H_2S		800	23	1104	283.1	0.2
H_2O (v)		268,900	14,926	1104	1,558.8	419.2
Sub-Total		3,402,400	143,417			1,447.4
Heat Exchange						112.5
Heat[1]						2.6
Totals		3,402,400	143,417			1,562.5
Low Temperature Side						
Input						
Gas	15					
CO_2		59,400	1,350	200	29.2	1.7
H_2O (v)		800	42	200	1,124.9	0.9
Sub-Total		60,200	1,392			2.6
Gas	16					
CO_2		777,200	17,659	676	143.9	111.8
H_2O (v)		341,700	18,967	676	1,343.9	459.2
Sub-Total		1,118,900	36,626			571.0
Heat Exchange						112.5
Totals		1,179,100	38,018			686.1
Output						
Gas	10					
CO_2		836,600	19,009	943	215.9	180.6
H_2O (v)		342,500	19,009	943	1,476.0	505.5
Totals		1,179,100	38,018			686.1

(1) By difference, to force the heat balance.

Table B-VII

Mass and Heat Balance
Heat Exchanger C-206
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	23	2,131,100	89,822	1200	-	978.7
Output						
Gas	13					
CH_4		31,000	1,930	1099	775.6	24.0
H_2		29,700	14,725	1099	3,625	107.7
CO		463,400	16,544	1099	269.6	124.9
CO_2		322,300	7,323	1099	259.9	83.8
N_2		1,110,900	39,652	1099	267.4	297.1
NH_3		4,900	285	1099	675.0	3.3
H_2S		500	15	1099	281.6	0.1
H_2O		168,400	9,348	1099	1,556.2	262.1
Sub-Total		2,131,100	89,822			903.0
Heat Exchange						75.7
Totals		2,131,100	89,822			978.7
Low Temperature Side:						
Input						
Gas from CO_2 Plant	26					
CH_4		6,100	377	230	97.5	0.6
H_2		5,800	2,875	230	585.1	3.4
CO		90,500	3,230	230	42.3	3.8
CO_2		3,500	80	230	35.8	0.1
N_2		216,900	7,743	230	42.3	9.2
NH_3		1,000	56	230	97.3	0.1
H_2S		100	3	230	41.3	-
H_2O		27,700	1,539	230	1,138.1	31.5
Sub-Total		351,600	15,903			48.7
Heat Exchange						75.7
Totals		351,600	15,903			124.4
Output						
Gas from CO_2 Plant	14					
CH_4		6,100	377	870		3.5
H_2		5,800	2,875	870		16.3
CO		90,500	3,230	870		18.8
CO_2		3,500	80	870		0.7
N_2		216,900	7,743	870		44.7
NH_3		1,000	56	870		0.5
H_2S		100	3	870		0.0
H_2O		27,700	1,539	870		39.9
Totals		351,600	15,903			124.4

Table B-VIII

Mass and Heat Balance
Heat Exchanger C-207
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas	13	2,131,100	89,822	1099	-	903.0
Output						
Gas	24					
CH_4		31,000	1,930	671	404.1	12.5
H_2		29,700	14,725	671	2,118	62.9
CO		463,400	16,544	671	154.7	71.7
CO_2		322,300	7,323	671	142.6	46.0
N_2		1,110,900	39,652	671	154.0	171.1
NH_3		4,900	285	671	371.0	1.8
H_2S		500	15	671	157.0	0.1
H_2O		168,400	9,348	671	1,341.5	225.9
Sub-Total		2,131,100	89,822			592.0
Heat Exchange						310.3
Heat Loss(1)						0.7
Totals		2,131,100	89,822			903.0
Low Temperature Side:						
Input						
Air						
O_2		511,800	15,994	398	73.9	37.8
N_2		1,686,200	60,186	398	84.4	142.3
H_2O (v)		6,000	329	398	1,213.5	7.3
Sub-Total		2,204,000	76,509			187.4
Heat Exchange						310.3
Totals		2,204,000	76,509			497.7
Output						
Air	31	2,204,000	76,509	946	-	497.7

(1) By difference, to force the heat balance.

Table B-IX

Mass and Heat Balance
Compressor JC-203
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Gas	24	2,131,100	89,822	671	-	592.0
Isentropic Work = $\bar{c}_p T_1 \left[(P_2/P_1)^{(N-1)/N} - 1\right]$ (89,822 mols) =						18.2(1)
Inefficiency = (18.2/0.89) - 18.2 =						2.2(2)
Totals		2,131,100	89,822			612.4
Output						
Recycle Gas	30					
CH_4		24,900	1,553	700	426.9	10.6
H_2		23,900	11,850	700	2,219	53.0
CO		372,900	13,314	700	162.3	60.5
CO_2		259,400	5,893	700	150.2	39.0
N_2		894,000	31,909	700	161.5	144.4
NH_3		3,900	229	700	390.4	1.5
H_2S		400	12	700	165.0	0.1
H_2O		135,500	7,523	700	1,355.5	183.7
Sub-Total		1,714,900	72,283			492.8
Gas to CO_2 Plant	32					
CH_4		6,100	377	700	426.9	2.6
H_2		5,800	2,875	700	2,219	12.9
CO		90,500	3,230	700	162.3	14.7
CO_2		62,900	1,430	700	150.2	9.4
N_2		216,900	7,743	700	161.5	35.0
NH_3		1,000	56	700	390.4	0.4
H_2S		100	3	700	165.0	-
H_2O		32,900	1,825	700	1,355.5	44.6
Sub-Total		416,200	17,539			119.6
Totals		2,131,100	89,822			612.4

(1) $\bar{c}_p$ = 7.832 Btu/lb mol °R
T_1 = 1131 °R
$N = \bar{c}_p/(\bar{c}_p - 1.99)$ = 1.3406
P_2/P_1 = 1.0927 ($\Delta P \sim 19$ psi)

(2) 89% efficiency

Table B-X

Mass and Heat Balance
Heat Exchanger C-302
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
High Temperature Side:						
Input						
Gas to CO_2 Plant	32	416,200	17,539	700	-	119.6
Output						
Gas to CO_2 Plant	25					
CH_4		6,100	377	236	101.1	0.6
H_2		5,800	2,875	236	605.9	3.5
CO		90,500	3,230	236	43.8	4.0
CO_2		62,900	1,430	236	37.1	2.3
N_2		216,900	7,743	236	43.7	9.5
NH_3		1,000	56	236	100.7	0.1
H_2S		100	3	236	42.8	-
H_2O		32,900	1,825	236	1,140.8	37.5
Sub-Total		416,200	17,539			57.5
Heat Exchange						62.1
Totals		416,200	17,539			119.6
Low Temperature Side:						
Input						
Boiler Feed Water	33	406,300	22,554	60	0	0.0
Heat Exchange						62.1
Totals		406,300	22,554			62.1
Output						
Boiler Feed Water	35	406,300	22,554	213	153	62.1

Table B-XI

Mass and Heat Balance
Product Gas at 206 psia
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

	Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input						
Gas	22	3,402,400	143,417	1104	-	1,447.4
Gas from CO_2 Plant	14	351,600	15,903	870	-	124.4
Totals		3,754,000	159,320			1,571.8
Output						
Product Gas	17					
CH_4		55,500	3,459	1081	758.5	42.1
H_2		53,200	26,387	1081	3,562	189.5
CO		830,400	29,646	1081	264.7	219.8
CO_2		518,100	11,773	1081	254.8	132.0
N_2		1,990,600	71,053	1081	262.5	522.5
NH_3		8,700	511	1081	661.4	5.8
H_2S		900	26	1081	276.1	0.2
H_2O (v)		296,600	16,465	1081	1,546.8	458.8
Sub-Total		3,754,000	159,320			1,570.7
Heat Loss (1)						1.1
Totals		3,754,000	159,320			1,571.8

(1) By difference, to force the heat balance.

Table B-XII

Mass and Heat Balance for Expander
Product Gas at 25.7 psia
High Gasification Case: See Figure B-1

Basis: 1 hour Datum: 60°F, H_2O (ℓ)

Stream	Pounds	Mols	Temp. °F	Enthalpy Δh Btu/lb	Enthalpy ΔH MM Btu
Input					
Product Gas at 206 psia	17	3,754,000	159,320	1081	-	1,570.7
Output					
Work = $-0.91\ \bar{c}_p\ T_1 \left[(P_2/P_1)^{(N-1)/N} - 1\right]$ (159,320 mols) =					721.3(1)
Product Gas at 25.7 psia					
CH_4	55,500	3,459	512	284.6	15.8
H_2	53,200	26,387	512	1,563	83.2
CO	830,400	29,646	512	113.6	94.3
CO_2	518,100	11,773	512	102.1	52.9
N_2	1,990,600	71,053	512	113.2	225.3
NH_3	8,700	511	512	267.7	2.3
H_2S	900	26	512	113.8	0.1
H_2O	296,600	16,465	512	1,266.1	375.5
Sub-Total	3,754,000	159,320			849.4
Totals	3,754,000	159,320			1,570.7

(1) 91% efficiency
$\bar{c}_p$ = 7957 Btu/lb mol °R
T_1 = 1541°R
$N = \bar{c}_p/(\bar{c}_p - 1.99) = 1.3335$; $(N-1)/N = 0.2501$
$P_2/P_1 = 25.7/206 = 0.1248$

Table B-XIII

Mass Balance
CO_2 Removal System
High Gasification Case: See Figure B-1

Basis: 1 hour

	Stream	Pounds	Mols	Temp. °F
Input				
Gas to CO_2 Plant	25			236
CH_4		6,100	377	
H_2		5,800	2,875	
CO		90,500	3,230	
CO_2		62,900	1,430	
N_2		216,900	7,743	
NH_3		1,000	56	
H_2S		100	3	
H_2O (v)		32,900	1,825	
Totals		416,200	17,539	
Output				
Gas from CO_2 Plant	26			230
CH_4		6,100	377	
H_2		5,800	2,875	
CO		90,500	3,230	
CO_2		3,500	80	
N_2		216,900	7,743	
NH_3		1,000	56	
H_2S		100	3	
H_2O (v)		27,700	1,539	
Sub-Total		351,600	15,903	
Make-up CO_2	15			200
CO_2		59,400	1,350	
H_2O (v)		800	42	
Sub-Total		60,200	1,392	
Make Water	20	4,400	244	-
Totals		416,200	17,539	

Table B-XIV

Mass Balance
Sulfur Recovery
High Gasification Case: See Figure B-1(1)

Basis: 1 hour

	Stream	Pounds	Mols	Temp. °F
Input				
Gas to Sulfur Recovery	12			1300
CO_2		777,200	17,659	
H_2S		46,000	1,350	
H_2O (v)		318,200	17,659	
Sub-Total		1,141,400	36,668	
Combustion Air	-			-
O_2		21,800	682	
N_2		71,900	2,566	
H_2O (v)		200	14	
Sub-Total		93,900	3,262	
Totals		1,235,300	-	
Output				
Gas	16			676
CO_2		777,200	17,659	
H_2O		341,700	18,967	
Sub-Total		1,118,900	36,626	
Vent Gas	18			90
SO_2		400	7	
N_2		71,900	2,566	
H_2O (v)		100	8	
Sub-Total		72,400	2,581	
Product Sulfur	19	43,100	1,343	
Make H_2O	-	900	48	
Totals		1,235,300	-	

(1) Small streams to and from the spent acceptor stripping column have been neglected.

ENVIRONMENTAL PROTECTION AGENCY

TECHNICAL PUBLICATIONS BRANCH
OFFICE OF ADMINISTRATION
RESEARCH TRIANGLE PARK, N.C. 27711

OFFICIAL BUSINESS
PENALTY FOR PRIVATE USE, $300

POSTAGE AND FEES PAID
ENVIRONMENTAL PROTECTION AGENCY
EPA - 335

RECEIVED
NATIONAL TECHNICAL
JAN 30 1973
INFORMATION SERVICE